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Fluidized-Bed Gasification of Peat, Lignite Subbituminous, and Pretreated Bituminous Coal

North Carolina State Univ. at Raleigh

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FLUIDIZED-BED GASIFICATION OF PEAT, LIGNITE, SUBBITUMINOUS, AND PRETREATED BITUMINOUS COAL

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16. ABSTRACT The report summarizes and compares results obtained from gasifying four different feedstocks in a pilot-scale fluidized-bed gasifier. Effects of operating variables (e.g., temperature, steam/carbon feed ratio, bed height, and feed rate) on carbon conversion and gas production are described. Gas compositions, expressed as mole percentages on a dry N2-free basis, showed only small differences, except for sulfur gases. No correlation of wastewater species production rates with reactor operating conditions was observed. Tar production rate was greatest for a New mexico subbituminous coal. A coastal peat and a Texas lignite gave similar tar production rates.

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ABSTRACT

A devolatilized Kentucky bituminous coal, a New Mexico subbituminous coal, a North Carolina peat, and a Texas lignite were gasified with steam and oxygen in a pilot-scale fluidized-bed reactor. The reactor was operated at pressures of 570-840 kPa (80-120 psia), molar steam-to-carbon feed ratios of 0.6 to 1.9, and average bed temperatures of 795-1010°C (1460-1850°F). The coal feed rate ranged from 14 to 33 kg/hr (30-73 lb/hr). All reactor effluent streams were measured and analyzed, enabling performance of mass and energy balances and identification of potentially hazardous species.

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INTRODUCTION

Rising prices and dwindling reserves of natural fuel sources, coupled with readily available reserves of coal, have made coal an increasingly attractive source of energy and chemical feedstocks. For these reasons, much research has centered on such coal conversion processes as pyrolysis, gasification, and liquefaction.

Most research has dealt with optimizing the output of coal conversion processes, while the environmental aspects have received considerably less attention. The U. S. Environmental Protection Agency (EPA), noting this gap in technology, began a study of the environmental consequences of large-scale coal conversion processes. As a part of this study, the EPA contracted for the construction of a coal gasification and gas cleaning pilot plant at North Carolina State University (NCSU). The facility was designed and built by Acurex-Aerotherm Corporation. Design of the gasification reactor, a fluidized bed unit, was subcontracted to the Institute of Gas Technology. Completed in 1978, the facility is now operated by the faculty, students, and staff of the NCSU Chemical Engineering Department.

The overall objective of the NCSU coal gasification project is to characterize the emissions from the gasification and gas cleaning processes, and to determine how the emission rates of pollutants depend on coal feedstocks and process operating conditions. While the overall project objectives include both gasification and gas cleaning, the research to be described in this report is primarily concerned with the gasification process.

In the sections that follow, results from the gasification of four feed materials, Kentucky devolatilized bituminous coal, New Mexico subbituminous coal, North Carolina peat, and Texas Lignite will be presented and discussed.

COAL GASIFICATION PROCESSES

Coal gasification technology has been in existence for many years. As early as the 18th century gas was produced by devolatilizing coal. The first coal gas company in England was chartered in 1812, and the first in the U.S. was chartered in 1816 (Perry, 1974a). The gas was used primarily in gas lighting.

Devolatilization leaves a large portion of the coal unconverted; thus methods to obtain greater conversions were developed. The methods developed produced a gas called "water gas" or "blue gas". The gas was produced in a cyclic scheme, consisting of alternating periods of partial combustion with air to heat the coal, and reaction with steam to produce carbon monoxide and hydrogen. The gas was enriched to a heating value of around 550 Btu/ft^3 by the "carburetted water gas" process developed by Lowe in 1873 (Lowry, 1945). Oils were cracked into lighter hydrocarbons on bricks heated by the combustion phase gas, and the resulting gas was blended with the steam phase gas. This process was inefficient, difficult to operate, and costly. A cheaper, simpler process produced a gas known as "producer gas" (Perry, 1974b). A continuous feed of both air and steam was used, and a product gas heating value of up to 150 Btu/ft^3 was obtained. "Carburetted water gas" and "producer gas" provided a large portion of the U.S. gas needs up into the 1930's. However, after World War II natural gas availability improved rapidly, and the use of coal gas steadily declined.

In recent years, interest in coal gasification has revived as future supplies of petroleum and natural gas are becoming less certain. The initial efforts in coal gasification utilized air-blown fixed bed reactors at low pressure. More recent work has centered on oxygen-blown reactors at high pressure. Air-blown reactors yield a lower heating value product gas due to the nitrogen in air. Three main classes of gasification reactors have now been developed: fixed bed, fluidized bed, and entrained bed.

Fixed bed reactors, sometimes more appropriately called moving bed reactors, employ a slowly moving bed where solids are fed from the top and residue is removed from the bottom. The countercurrent feed gas rate is such that the solids are not stirred, and no significant mixing occurs. Fixed beds generally attain high carbon conversions due to the plug flow of the solids and relatively long residence times, high thermal efficiencies, and small solids losses by entrainment. However,

fixed beds often cannot handle caking coals; hot spots can develop due to the lack of mixing; and hydrocarbon and tar emissions are generally greater than those from other reactor types.

The fixed bed is the most frequently used commercial reactor type, with the Lurgi process (Rudolph, 1973) being the best known example. The Lurgi gasifier is a pressurized steam-oxygen gasifier, is nonslagging, and typically operates at 300-400 psia and 1000-1600 ^{o}F. After shift conversion and methanation, a 970 Btu/SCF gas is obtained. The SASOL plant in South Africa, the largest gasification plant currently in operation, uses this gasifier. Other fixed bed gasifiers which have been used commercially include Wellman-Galusha, Willputte, ATC/Wellman, and Woodall-Duckham. A summary of fixed bed gasification operations taken from Wett (1980) and Elliott (1981) is given in Table 1.

TABLE 1
FIXED BED GASIFIERS

Process	Pressure	Oxidant	Agglomeration Prevention
Dry Ash, 1-Stage			
Gegas	to 500 psig	air	stirrer paddles
Kellogg	10 psig	air or O_2	stirrer
Lurgi*	to 450 psig	air or O_2	rotating blades
Merc	to 105 psig	air	spiraling stirrer
Riley-Morgan	40 in. H_2O	air	agitator in rotating bed
Wellman-Galusha*	atmospheric	air	spiraling arms
Wilputte*	atmospheric	air	rotating arm
Dry Ash, 2-Stage			
ATC/Wellman	atmospheric	air	none
Foster-Wheeler/Stoic*	atmospheric	air	none
Ruhr-100	1500 psig	oxygen	stirrer blades
Woodall-Duckham*	atmospheric	air or O_2	none
Slagging			
BCURA/CJG*	45 psig	oxygen	---
BGC/Lurgi*	to 400 psig	oxygen	stirrer
Gferc	to 400 psig	oxygen	stirrer

* Commercially used

In entrained bed gasifiers the solids are completely suspended in a high velocity cocurrent gas flow. High temperatures, generally over 2000°F, give rise to high reaction rates and carbon conversions approaching 100%. The solids residence time is small, so that high throughput rates can be achieved. The high operating temperature generally eliminates tar and heavy hydrocarbon emissions, which are cracked to form lighter, less harmful species. In addition, practically any feed material can be used, as high turbulence prevents agglomeration. Drawbacks to entrained bed gasification include the requirement for high temperature materials, difficulties in handling the molten ash, and high oxygen requirements. The high oxygen rate and the short solids residence time could cause problems if the solids feed were interrupted; thus good instrumentation is also critical.

The Koppers-Totzek atmospheric gasifier is probably the most thoroughly tested entrained bed gasifier. It is a low pressure, dry feed, steam-oxygen gasifier, with operating temperatures over 2700°F (LeBlanc, 1981; Staege, 1979). The most promising new entrained gasification systems are the Shell and Texaco processes. The Shell process is basically a pressurized modification of the Koppers-Totzek process. The Texaco gasifier differs in that it utilizes a coal-water slurry feed, downflow of reactants, and higher pressures. Table 2, compiled from the data of Wett (1980) and Elliott (1981), gives a summary of entrained bed gasifiers.

TABLE 2
ENTRAINED BED GASIFIERS

Process	Pressure	Gas Feed	
1 Stage			
Babcock-Wilcox	300 psig	oxygen or air, and steam	
Bell HMF	to 225 psig	air and steam	
Koppers-Totzek*	10-12 psig	oxygen and steam	
Mountain fuel	to 150 psig	oxygen and steam	
Saarberg/Otto	to 360 psig	oxygen or air, and steam	
Shell	to 500 psig	oxygen or air, and steam	
Texaco	to 1200 psig	oxygen or air, and steam	
2 Stage		1st Stage	2nd Stage
Bi-gas	to 1500 psig	syngas and steam	oxygen and steam
Combustion Eng.	atmospheric	combustion gas and steam	air
Foster Wheeler	atmospheric	syngas and steam	air and steam
Peatgas	to 500 psig	syngas	oxygen and steam
Rockwell Int.	to 1500 psig	hydrogen	unknown

* Commercially used

Fluidized beds have characteristics between those of fixed and entrained beds. The fluidizing gas is fed into the bottom of the reactor while solids may be fed from the top or bottom. When the gas rate is sufficient, perfect mixing of the solids is approached; however, slugging and excessive entrainment may occur when the gas rate is too great. Top solids feed tends to produce more tars and heavy hydrocarbons as devolatilization occurs above the bed and the products are swept away before cracking reactions may occur. In bottom feed systems, devolatilization products have longer residence times and may undergo cracking to a greater extent. Among the desirable traits of fluidized beds are that high heat and mass transfer rates are achieved; operation is nearly isothermal due to the rapid mixing; additives, such as catalysts or limestone, are easily mixed with the coal feedstock; small particles, giving high surface to volume ratios, may be used; and control is simplified by the uniform conditions.

The Winkler gasifier (Wall, 1982) is the only fluidized bed system which has seen extensive commercial use. It is an atmospheric reactor, and is operated with air or oxygen at 1500-2000^{o}F. In 1979 there were 70 Winkler gasifiers in operation at 24 installations. Many other fluidized bed gasifiers are being developed. Table 3 lists several fluidized bed gasification systems (Wett, 1980; Elliott, 1981).

TABLE 3
FLUIDIZED BED GASIFIERS

Process	Pressure	Oxidant
Battelle-Union Carbide	100 psig	air
BCR Tri-Gas	250 psig	air
British Gas	750 psig	oxygen
CO_2 Acceptor	150 psig	air
Cogas	10 psig	air
Exxon Catalytic	500 psig	none
Hydrane	1000 psig	oxygen
Hygas	1200 psig	oxygen or air
Peatgas	500 psig	oxygen
Rheinbraun	150 psig	oxygen
Synthane	1000 psig	oxygen
U-Gas	350 psig	oxygen or air
Westinghouse	225 psig	air
Winkler*	atmospheric	oxygen or air

* Commercially used

PILOT PLANT FACILITY

A complete description of the pilot plant facility is given by Ferrell et al., (1980). The plant consists of the following major subsystems: gasifier, raw gas cleaning, and acid gas removal. The following sections provide a description of each subsystem.

GASIFIER

The reactor, shown in Figure 1, is a 15.2 cm (6 in.) I.D. Schedule 40 pipe (316 SS) enclosed in several layers of Fiberfrax bulk ceramic insulation and contained in a 61 cm (24 in.) I.D. Schedule 80 carbon steel pipe. The overall height of the unit is roughly 3.7 m (12 ft). The gas feed is introduced into the reactor through three feed nozzles spaced triangularly near the bottom of the reaction chamber. Each nozzle consists of a 0.48 cm (3/16 in) orifice inside a diverging cone. Coal is fed at the top of the reactor and removed at the bottom by nitrogen-purged screw conveyors. The feed hopper is a 51 cm (20 in) I.D. schedule 80 carbon steel pipe, with a volume of 0.6 m^3 (22 ft^3). The char receiver is a 61 cm (24 in) I.D. schedule 80 carbon steel pipe with a volume of 0.5 m^3 (18 ft^3).

The temperature profile in the bed is monitored by means of six thermocouples in a central thermowell in the reactor. The thermocouples are located at positions 13, 25, 64, 89, 114, and 140 cm (5,10,25,35,45,55 in) above the feed nozzles. The 25 cm thermocouple is used for reactor temperature control, with the oxygen feed rate being adjusted to maintain the selected temperature. A reactor pressure tap is located at the top of the reactor. Differential pressure taps on the gas feed line below the feed nozzles, at 38 cm and 89 cm above the feed nozzles, and at the top of the reactor are used to monitor the pressure drop across the feed cones and two pressure drops in the bed. The level of the fluidized bed is monitored with a nuclear level gauge, and is controlled by adjusting the char removal screw rotation rate. The height of the bed above the feed nozzles is normally 97 cm (38 in), although occasional runs with a 132 cm (52 in) bed have been made.

Nitrogen is supplied to the gasifier by a 23 m^3 (6000 gallon) liquid nitrogen tank and vaporizer system. Oxygen is fed from a manifold of six cylinders, and steam feed is from a 1135 kPa (150 psig) building process steam line.

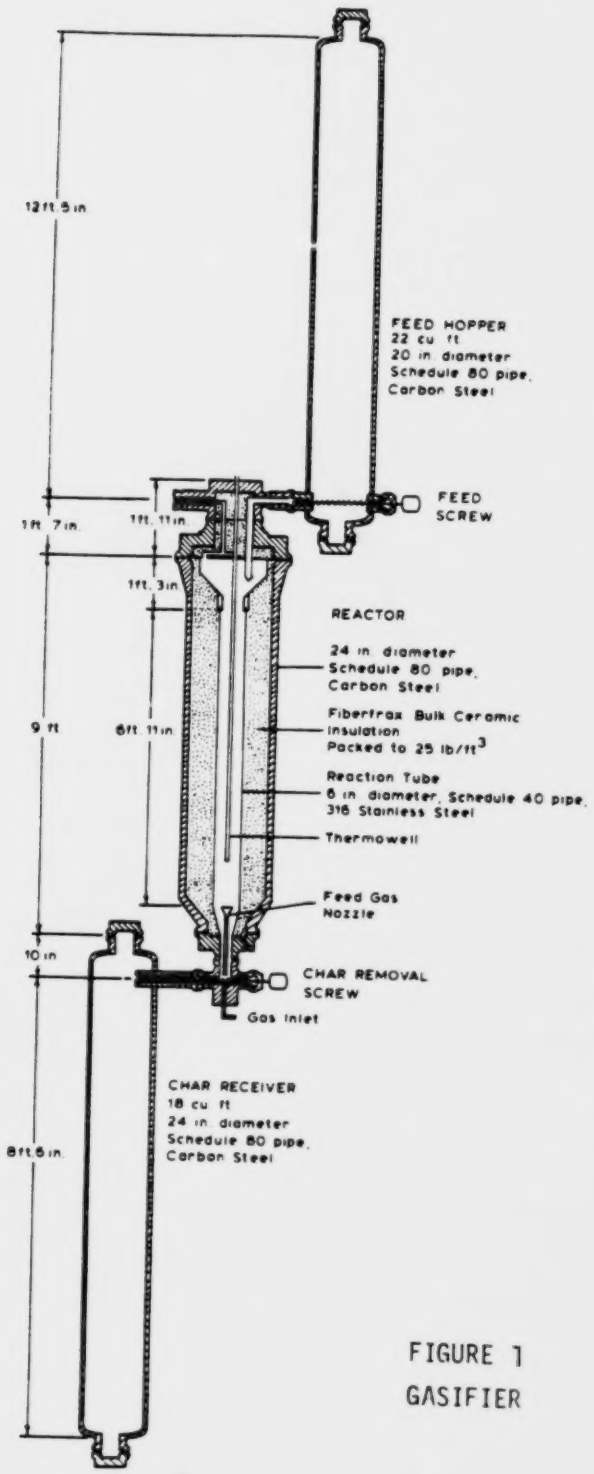

FIGURE 1
GASIFIER

PCS SYSTEM

The PCS (particulates, condensables, and solubles removal) system is shown in Figure 2. The gas exiting the gasifier is passed through a cyclone separator, where most of the elutriated particles are removed and collected in a fines collection vessel. After the cyclone, the gas line expands to 7.6 cm (3 in), and the gases pass through a scrubber system. The scrubber system consists of a set of three spray nozzles to quench the gas and a reduction of the pipe diameter to 2.5 cm (1 in) to promote gas-liquid contacting. The scrubber water is pumped by a 2.5 l/min (40 GPH) positive displacement pump, and may be either recirculated condensate or fresh city water.

The condensate accumulates in the PCS tank, a 2.5 m (8 ft) vessel with 0.9 m (3 ft) O.D. In the PCS tank a set of cooling coils -- three 6.1 m (20 ft) sections of 1.3 cm (1/2 in) tubing with city water as the coolant -- maintain the condensate temperature well below its boiling point. The emerging gas passes through a single-pass shell-and-tube heat exchanger, with city water on the shell side and the gases on the tube side. The exchanger is situated vertically above the PCS tank so that condensing water may run back into the tank. A demister, water separator, and coalescing filter are then used to remove any remaining water. The gas leaving the filter is either burned in a shielded flare or fed to the acid gas removal system.

Differential pressure taps measure the pressure drops across the cyclone, spray nozzles, and coalescing filter. The temperature is measured by thermocouples at five points in the PCS system. An orifice meter downstream of the filter is used to measure the make gas flow rate.

ACID GAS REMOVAL SYSTEM

The acid gas removal system (AGRS) is shown in Figure 3. Although not specifically a part of this research, the AGRS is a major portion of the pilot plant facility and thus will be described briefly. Kelly (1981) and Ferrell et al. (1980) provide detailed descriptions of the AGRS.

The major components of the AGRS are an absorption column, a flash tank, and a stripping column with the necessary auxiliary equipment. The major purpose of the system is to remove carbon dioxide and hydrogen sulfide from the gasifier make gas. In so doing it also usually removes most of the other sulfur gases and many of the hydrocarbons of molecular weight higher than methane. The system is modular in design so that alternative absorption processes can be evaluated. It was designed to

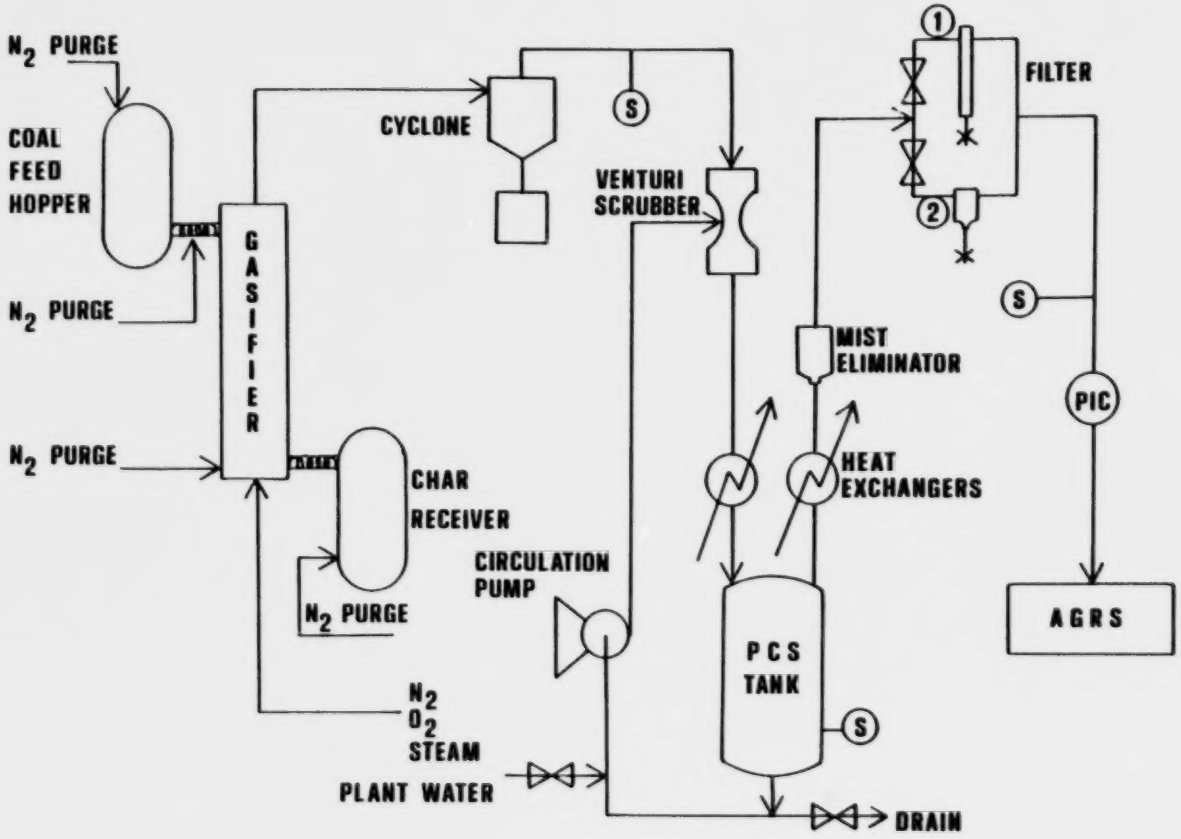

FIGURE 2

PCS SYSTEM

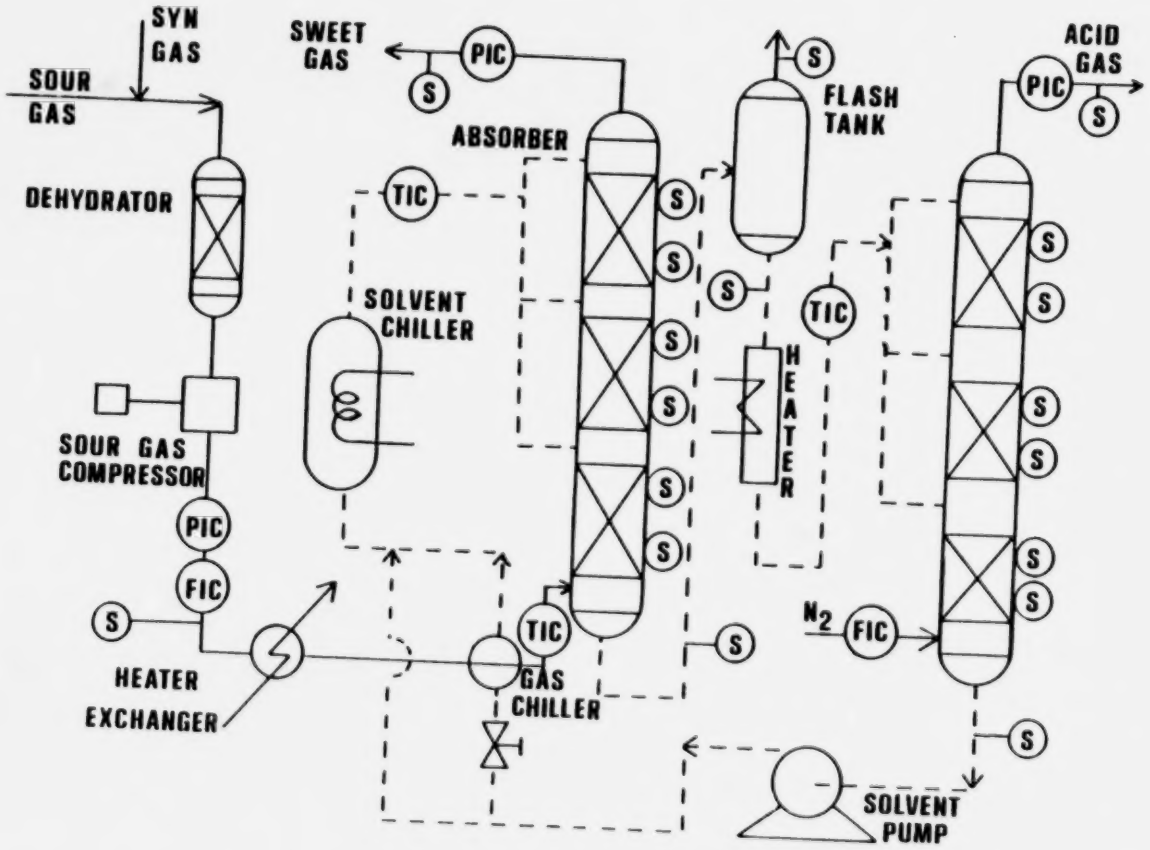

FIGURE 3

ACID GAS REMOVAL SYSTEM

operate with four different solvents; however, refrigerated methanol has been the only solvent used so far.

Gases from the PCS system are fed through a pair of dehydrating towers, compressed, cooled, and fed to the absorber. The absorption column contains three sections of ceramic Intalox saddle packing. The column is well-insulated, and approaches adiabatic operation. The solvent exiting the absorber is fed to a flash tank where some of the absorbed gases are released. The solvent then exchanges heat with the solvent exiting the stripper and is fed to the stripper. The stripper is similar to the absorber in design in that it contains three packed sections of ceramic Intalox saddles. The column approaches adiabatic operation, and the solvent can be regenerated either by stripping with an inert gas or with reboiled methanol vapors.

DATA ACQUISITION AND PROCESS CONTROL

Approximately 100 process sensors are used to obtain information during pilot plant operation. The plant is controlled with a Honeywell TDC-2000 process control computer which has a capability of regulating 16 process control loops throughout the pilot plant. Also, an alarm system incorporating all process sensors calls attention to process problems by sounding an audible alarm and lighting an alarm light on a process schematic above the control panel. The alarm system can also be set to cause plant shutdown automatically at pre-set alarm points.

The 100 process sensors (temperatures, pressures, differential pressures, flow rates, and levels) are monitored through a data acquisition system linked to a PDP 11/34 minicomputer. Each sensor is interpreted every five seconds for display on a video terminal. The information is averaged and stored every five minutes on magnetic disk for future data reduction. The capability of plotting information from each process sensor for the previous hour on a video or hard-copy terminal during the run also exists. Strip chart recorders are used for continuous monitoring of process temperatures and selected process variables.

Willis (1981) gives a detailed description of the process control and data acquisition system.

PILOT PLANT OPERATION

STARTUP AND STEADY-STATE OPERATION

The coal feed hopper is loaded the day before a run. Depending on the desired length of the run, 230-400 kg (500-900 lb) of coal are loaded. Up to 45 kg (100 lb) may be used in startup, and roughly 70 kg (150 lb) must be in the feed hopper to insure that the hopper does not run dry. This leaves around 110-300 kg (250-650 lb) for steady-state operation. At a normal feed rate of 23 kg/hr (50 lb/hr), 5-13 hours of steady-state operation can thus be achieved. For runs not utilizing the AGRS, five hours is more than sufficient to insure steady operation and obtain multiple sets of samples. When the AGRS is used a longer run is required, as the AGRS steady-state is generally reached 2-3 hours after that of the gasifier. In addition, 7 kg (15 lb) of a devolatilized coal (char) is loaded into the reactor to aid in startup.

After the feed hopper is loaded, the system is closed and a pressure test is run. This is done to insure that no major leaks are present, and to estimate the leak rate for later use in the mass balance. The gasifier-PCS system is pressurized with nitrogen to operating pressure, all valves are closed, and the system pressure is monitored. Assuming ideal gas behavior and constant temperature, the leak rate is estimated as

$$\text{Leak Rate} = \frac{dn}{dt} = \frac{V}{RT}\frac{dP}{dt} \qquad (1)$$

The system volume V is calculated from the known empty system volume minus the volume of the coal in the feed hopper and the water in the PCS tank. The pressure decay rate, dP/dt, is determined by a least-squares fit of pressure vs time data taken over a 15 minute span. The estimated leak rate is normally 0.015 scmm (0.5 scfm) or less. A considerable portion of the leak rate is thought to be internal, possibly due to valve leakage, as no significant external leaks can be detected.

The system is started up by blowing nitrogen preheated to 540°C (1000°F) through a shallow char bed. The bed is allowed to heat to 260°C (500°F), at which time a small coal feed is started. The coal feed aids in bed ignition. When the bed temperature measured by the 13 cm (5 in) thermocouple reaches 320-370°C (600-700°F) oxygen feed is

started. The bed temperature then rises rapidly as oxygen flow is increased and nitrogen flow is decreased.

The bed, composed mostly of char, burns relatively cleanly and very little tar formation occurs. Startup with a pure coal bed would most likely produce large amounts of tar that would condense in the unheated reactor exit line, forcing an early shutdown. Startup with char heats the system so that when the coal feed is increased to the desired steady-state value, tar condensation is not a problem.

Steam preheated to 540°C (1000°F) is fed at about 7 kg/hr (15 lb/hr), beginning when the bed temperature reaches 650-760°C (1200-1400°F). The steam feed is slowly increased to the selected steady-state value, while oxygen is increased as necessary to bring the bed temperature to the desired level. As steam and oxygen flows are increased, nitrogen flow is simultaneously decreased to maintain the inlet gas velocity near 0.3 m/s (1 ft/s).

While the inlet gas rates are being brought to their steady-state values, the coal feed rate is increased to its final value. The fluidized bed level increases until it is detected by the nuclear level gauge. The level is then controlled by adjusting the rate of the char removal screw.

The process control and data acquisition system allows all gasifier-PCS system temperatures, pressures, differential pressures, levels, and flow rates to be monitored continuously. This allows the approach to steady-state to be closely observed. In addition, gas samples are taken and analyzed approximately every 30 minutes as steady-state operation is approached. Thus, the start of steady-state operation and its maintenance can be determined with good accuracy. The entire startup process generally requires about 4-5 hours.

The superficial velocity in the bed at steady state is on the order of 0.3 m/s (1.1 ft/s), which is three to five times the minimum fluidization velocity at the reactor conditions. There is a good likelihood that the bed is in a slugging mode of operation at steady state, but any consequent oscillations are too small to be detected by the system instrumentation. Energy balances on the reactor indicate that operation is essentially adiabatic.

SAMPLING

For each run, samples of all effluent streams are taken. Feed coal samples are taken the day before a run when the feed hopper is loaded. All other samples are taken either during the steady state portion of the run, or at the end of the run. The sample train, a sidestream

designed specifically for sampling purposes, and sampling techniques are described in the following sections.

Sample Train

A schematic diagram of the sample train is shown in Figure 4. For mass balances, the gas flow through the sample train is added to that metered at the PCS system outlet. During steady state operation, the gas flow through the sample train is metered with a test meter, and the water condensed in the train is measured and used as the primary measurement of the water content of the reactor make gas. Gas samples are taken in both stainless steel and glass bombs for GC analysis.

The gas sample train is designed to allow representative samples of particulates, tars, and condensed liquids to be taken. It is also necessary, especially in the case of trace element analysis, to trap all of the particulates, tars, and condensed liquids in one location to minimize sample contamination. For this purpose, a cold trap, which consists of a vertical stainless steel cylinder, equipped with gas inlet and exit ports, and a drain-plug to drain the condensed liquids, is used. The cylinder is loosely packed with stainless steel wool, which traps the particulates and tars as the gas travels through it. The entire trap is immersed in an ice bath. When the gasifier reaches steady-state, the cold trap is opened and the sample train gas is diverted through it for one hour. At the end of each run, the contents of the trap are separated into solid, tar, and liquid fractions, and sent for analysis.

Feed Coal

Multiple samples are taken the day before a run when the feed coal hopper is loaded. For each run proximate and ultimate analyses are performed. Sieve analyses are ocassionally done. Trace element analyses are also performed.

Spent Char

Multiple samples are taken from the char receiver vessel or from the bottom of the reactor after a run. Analyses are essentially identical to those for the feed coal. The steady state char removal rate is determined from the total char removed, the total revolutions of the removal screw, and the rate of revolution of the screw during the steady state period of operation.

Cyclone Fines

Samples are collected from the cyclone solids collection tank. Since the cyclone fines are sometimes wet, it is possible that some condensation on the fines occurs during or after shutdown. In addition, for several of the runs, very few solids are collected in the cyclone.

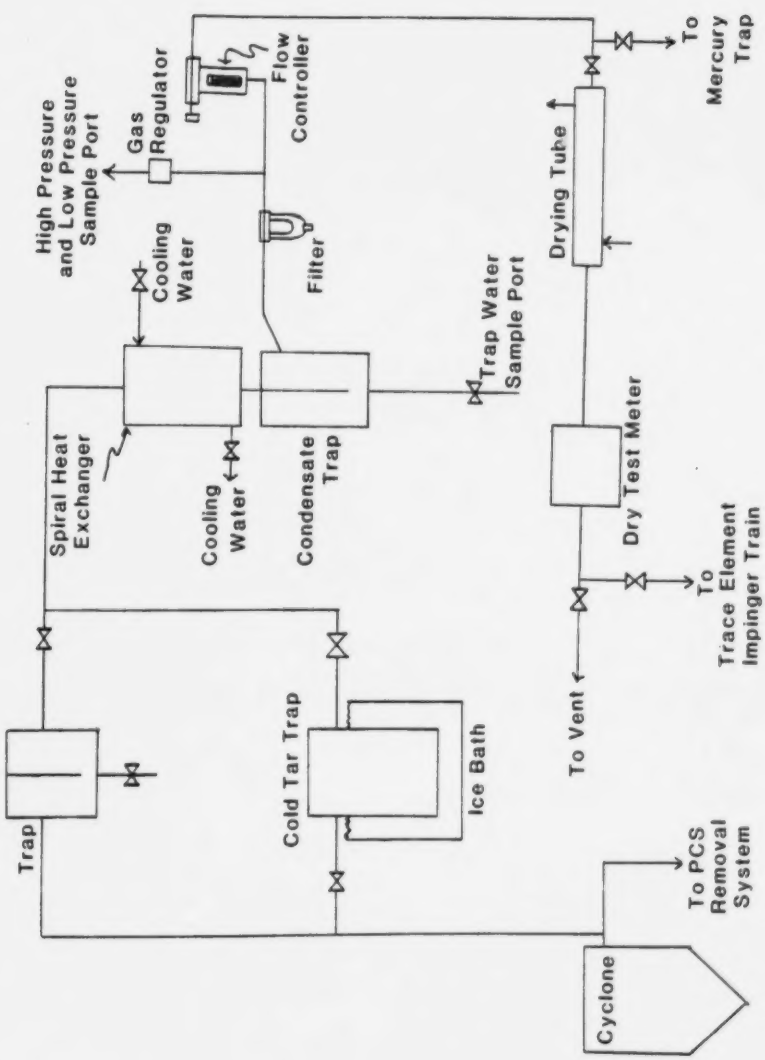

FIGURE 4
SAMPLE TRAIN

For these reasons, cyclone solids analyses are generally less reliable than those for the other solids. Analytical methods are the same as those for the feed coal.

PCS Tank Samples

After shutdown, samples of particulates, tar, and water are taken from the PCS tank. For all runs reported on here, the venturi scrubber water was recirculated from the tank to the spray nozzles, and thus contained material from the startup, steady state, and shutdown periods. Prior to each run, the tank is filled with approximately 600 liters (160 gallons) of fresh water, and thus the final composition of the water depends on the length of the run. Methods of analysis and results are presented later in this report.

Gas Samples

Gas samples are taken from the six locations noted on Figures 2 and 3 at operating pressure using one liter stainless steel bombs, and at lower pressures using one liter glass bombs. The glass bombs are coated internally with hexamethyldisilizane to reduce the adsorption of trace sulfur, nitrogen and hydrocarbon species.

Gases from the PCS system are metered and sampled after the final filter and before the dehydrators. Samples are collected at high pressure in stainless steel bombs, and at lower pressures -- approximately 240 kPa (20 psig) -- in glass bombs.

ANALYTICAL PROCEDURES

Solids

Both proximate and ultimate analyses are performed on all solid input and output streams. The proximate analyses are performed using a Fisher Coal Analyzer Model 490. A Lindberg Combustion Furnace Model 123-SP is used to determine carbon and hydrogen for the ultimate analyses, and a Fisher Sulfur Analyzer Model 470 is used for sulfur analysis. Nitrogen is determined by the Kjeldahl-Gunning method, and oxygen is determined by difference. Certified coal standards are analyzed periodically to check the accuracy of all analyses.

Liquids

The methods used for the routine analysis of the wastewater from the PCS tank and the sample train are listed below:

1. Ammonia - Acidimetric method 418D (Franson 1975). The sample pH is adjusted; the sample is distilled into boric acid, and the condensate is titrated with 0.1N H_2SO_4.

2. Cyanate - Pretreatment and heating by method 413J (Franson 1975), distillation and titration by method 418 (Franson 1975).

3. Carbon, TOC and TVC (Total organic carbon and total volatile carbon) - Samples are analyzed by a Dohrmann Envirotech Organic Carbon Analyzer, after being preserved by adjusting the pH to 2.0 with HCL. Thirty uL injections of the sample and potassium hydrogen phthalate standards are used.

4. Chloride, Fluoride, Sulfate, Sulfite - Analysis is by a Dionex System 10 Ion Chromatograph. Identification is by retention time comparison with standards. Concentrations are determined by calibration curves (micro-moles versus peak height) of standards.

5. COD (Chemical oxygen demand) - Analysis is by the Hach COD System, a compact system in which the dichromate reflux takes place in a small sealed vial. A titration determines the unreacted dichromate, and gives an estimate of the biological oxygen demand (BOD).

6. Nitrogen - Samples are analyzed by method 421 (Franson 1975). The sample is digested to convert the organic nitrogen to ammonia (Kjeldahl), and is then distilled and titrated as in the ammonia determination.

7. pH - Samples are determined by an Orion Microprocessor Ionalyzer 901. Calibration is with Fisher buffer solutions, and pH is determined with an Orion combination electrode.

8. Thiocyanate - Analysis is an adaptation of the method given by Franson (1975). A colorimetric method based on the formation of ferric thiocyanate at low pH is used. A sample blank is used instead of distilled water to remove interferences in the sample. Negation of preexisting ferric thiocyanate is accomplished by the addition of a small amount of sodium citrate to the sample blank.

9. Sulfide - Analysis is given by Luthy (1977). Sulfide is separated from solution by precipitation as zinc sulfide accompanied by filtration. The precipitate is redissolved in a known acid-iodine solution. Sulfide is oxidized by iodine, and back titrated with thiosulfate solution to the absence of blue color.

10. Phenolics - Analysis is done on reverse phase HPLC.

Water-methanol gradient, 12" stainless-steel C_{18} column, and UV detector at 280nm are used. Further details of sample preparation and analysis methods are given by Hubik and Ferrell (1982).

Gases

As noted earlier, gas samples are taken from the six locations noted on Figures 2 ane 3 at operating pressue using one liter stainless steel bombs, and at lower pressures using one liter glass bombs. Hydrogen, nitrogen, oxygen, carbon monoxide, carbon dioxide, methane, ethane, ethylene, and hydrogen sulfide (in relatively high concentrations) are analyzed on a Carle Series SX AGC-111H gas chromatograph in conjunction with a Perkin-Elmer Sigma 10 Chromatography Data System. The AGC-111H is a dual valve, dual column, dual detector unit. The first column is 6' x 1/8" 80% Poropak N and 20% Poropak Q. The second column is 7' x 1/4" Molecular Sieve 13X. A third column, 8% OV101 Chromasorb WAWDMCS, is used as a reference column. Columns are maintained at 75°C throughout the analyses.

Detection of all the abovementioned gases except hydrogen is performed with a thermal conductivity (TC) detector with helium carrier gas. Hydrogen is separated from the other gases by permeation through a palladium-tube transfer system. A second TC detector, with nitrogen carrier gas, detects H_2 at concentrations from 0.03% to 100%. The currents in both detectors are set at 20 mA.

Hydrogen sulfide, carbonyl sulfide, methyl mercaptan, ethyl mercaptan, carbon disulfide, and thiophene are analyzed on a Varian gas chromatograph and a Varian dual-flame flame photometric detector. A gas sampling valve is added for sample injection. Sample components are separated on a temperature-programmed 1.8m x 3mm (6' x 1/8") Teflon 1.5% x E-60, 1% H_3PO_4 Carbopak B column. The column is held at 25°C for 3 minutes, heated to 130°C at 15°C per minute, and held at 130°C for 6 minutes. The detector is heated to 180°C, and uses helium as carrier gas and hydrogen and air as flame gases.

Ethane, ethylene, all C_3 and C_4 hydrocarbons, and methanol are analyzed on a Tracor 550 gas chromatograph equipped with a flame ionization detector. A gas sampling valve and valve oven are added for sample injection. The sample components are separated on an 2.4m x 3mm (8' x 1/8") stainless steel Poropak Q column in series with a 25cm x 3mm (10" x 1/8") stainless steel 20% Carbowax 20M TPA column. The columns are temperature programmed from 50°C to 200°C at 8°C per minute. The flame ionization detector is heated to 250°C. Helium is used as the carrier gas.

Benzene, toluene, and xylenes are analyzed on the same gas chromatograph, using a 2.1m x 3mm (7' x 1/8") stainless steel 35% BC 150 Chromasorb PAW, DMCS column. The column is temperature programmed from 120°C to 200°C at 8°C per minute. The flame ionization detector is heated to 250°C. Helium is used as the carrier gas.

For all the gas chromatographic analyses, the Perkin-Elmer Sigma 10 Chromatography Data Station performs automated valve switching, receives all detector outputs, and integrates to calculate peak areas. The peak areas are converted to micromoles and mole percents by external standard calibration curves.

EXPERIMENTAL RESULTS

GASIFIER FEED MATERIALS

Four feed materials have been gasified in the NCSU pilot plant facility. The feeds were:

- o Kentucky devolatilized bituminous coal
- o New Mexico subbituminous coal
- o North Carolina peat
- o Texas lignite

A Western Kentucky #11 bituminous coal, devolatilized at 2000^{o}F in an inert atmosphere, was the initial feedstock. The pretreated coal, or char, was used since it was low in tar-forming volatiles and provided an easily gasified material for use in facility shakedown runs. The char was obtained from the Peabody Coal Company, Columbia, TN.

The New Mexico subbituminous coal used in pilot plant runs was obtained from the Utah International Company, Fruitland, NM. It was mined from the Navaho mine, then shipped to Morgantown, WV where it was ground and sieved. The coal was then fed to the gasifier with no further preparation.

The North Carolina peat was obtained from First Colony Farms of Creswell, NC. The peat was taken in milled (shredded) form, at roughly 50% moisture. Before being run in the NCSU pilot plant, it was screened to pass through a 1/4" screen, and was air-dryed to approximately 25% moisture.

The Texas lignite was received from Phillips Coal Company. The lignite was mined from a test pit east of Dallas, and shipped to NCSU in 3-inch chunk form at about 50% moisture. It was ground and sieved at NCSU, and air-dryed to roughly 25% moisture before running.

The four materials used in gasifier runs represented a wide range of coals. Several analyses were used to characterize the feeds. Proximate and ultimate analyses were routinely performed, while other analyses were done for selected runs. Analysis results are summarized in the following sections.

Proximate and Ultimate Analyses

Proximate and ultimate feed analyses were performed for all gasifier runs. Typical results for the four feed materials are given in Table 4. Complete results for all are given by Purdy (1982).

TABLE 4
GASIFIER FEED ANALYSES

	KY Char	NM Coal	NC Peat	TX Lignite
Proximate (wt %)				
Moisture	0.9	10.5	22.8	24.3
Vol. Matter	2.4	31.7	46.3	29.1
Ash	10.7	22.6	4.6	23.5
Fixed Carbon	86.0	35.2	26.3	23.1
Ultimate (wt %)				
Carbon	83.8	52.5	45.9	39.2
Hydrogen	0.6	4.8	4.3	4.2
Oxygen	2.2	18.1	44.1	32.1
Nitrogen	0.1	1.2	0.9	0.5
Sulfur	2.6	0.8	0.2	0.5
Ash	10.7	22.6	4.6	23.5

Sulfur Forms

The ultimate analysis provides the total mass fraction of sulfur in coal, but yields no information on the sulfur form distribution. The various sulfur forms behave differently during devolatilization and gasification, so that knowledge of the sulfur distribution is important in the analysis of sulfur emissions. Sulfur distributions for the Western Kentucky char (McCue, 1979), New Mexico coal (Agreda et al., 1979), and a North Carolina peat (Cleland et al., 1979) are given in Table 5. Data for the Texas lignite are unavailable.

TABLE 5
SULFUR FORM DISTRIBUTIONS

	Wt. Fraction of Total Sulfur		
	KY Char	NM Coal	NC Peat
Pyrite	0.047	0.425	0.083
Organic	0.949	0.575	0.500
Sulfate	0.004	0	0.417

Sieve Analyses

Sieve analyses were used to determine the size distributions of the gasifier feeds. U.S. Standard Sieves supplied by the Fisher Scientific Company were used. The sieve sizes used for all analyses were 12, 20, 30, 40, 60, 80, 100, 200, 325, and pan, the latter representing everything that passes through the 325 sieve. It is assumed that all of the solid sample passes through the 10 sieve, and that the material retained on any sieve tray has a diameter equal to the average of the opening of that tray and that of the next largest tray. The opening size and average diameter of retained particles are listed below for each sieve size.

Sieve	Opening Size, um	Avg. Particle Diameter, um
10	2000	---
12	1700	1850.0
20	850	1275.0
30	600	725.0
40	425	512.5
60	250	337.5
80	180	215.0
100	150	165.0
200	75	112.5
325	45	60.0
Pan	0	22.5

Overall average particle diameters were calculated by summing each weight fraction multiplied by its respective average particle size. An average feed size distribution was used for all char runs.

Density

Particle density and settled bed density were measured for several feed samples. Particle densities were determined by water displacement. Measurements of particle density were difficult for the peat and lignite due to the fines present in the materials and their high moisture adsorbency. For these two feeds, the particle density was estimated. Table 6 lists the results of density measurements.

TABLE 6
PARTICLE AND SETTLED BED DENSITIES

Sample	Densities, lb/ft^3 Settled Bed	Particle
Western Kentucky Char	46.2	112.2
	41.8	92.5
New Mexico Coal	51.8	82.6
	45.2	85.6
	48.0	
North Carolina Peat	35.4	75.0*
Texas Lignite	46.4	
	49.2	85.0*

* estimated

Ash Fusion

Ash fusion data were supplied by Phillips Petroleum for the Texas lignite. The results are given in Table 7 for both reducing and oxidizing environments.

TABLE 7
ASH FUSION TEMPERATURES FOR TEXAS LIGNITE

	Reducing atmosphere	Oxidizing atmosphere
Initial deformation	2227°F	2234°F
Softening temperature	2295	2306
Hemispherical temperature	2349	2360
Fluid temperature	2489	2495

EXPERIMENTAL RUNS

Five sets of runs were completed on the NCSU pilot plant coal gasification facility. The run sets were:

1. Kentucky devolatilized bituminous coal (char)
2. Mixtures of Kentucky char and New Mexico subbituminous coal
3. New Mexico subbituminous coal
4. North Carolina peat
5. Texas lignite

Sampling and analyses were performed for each run as described previously. Temperature and molar steam-to-carbon feed ratio were the main adjustable operating variables. The gasifier operating variables were as follows:

- o Average bed temperature - as low as 1475°F for lignite, up to 1900°F for char.
- o Pressure - normally 790 kPa (115 psia), although a few runs were carried out at pressures as low as 550 kPa (80 psia).
- o Solids feed rate - approximately 23 kg/hr (50 lb/hr), higher for high-moisture feeds. Due to variability of the feed screw efficiency, deviations from the desired rate occurred in some runs.

- o Steam feed rate - determined from solids feed rate and desired molar steam-to-carbon feed ratio, which was varied from 0.6 to 2.0.

- o Oxygen feed rate - as needed to maintain desired reactor temperature.

- o Nitrogen feed rate - as needed to maintain an inlet superficial gas velocity of roughly 0.3 m/s (1.1 ft/s).

- o Feed gas inlet temperature - 1000°F for all runs

- o Fluidized bed height - 38 in normally, although several char runs were at 52 in, and lower bed heights were used in some peat and lignite runs.

The runs and run conditions are discussed in the following sections. Several replicate runs were carried out to provide a consistent gas feed to the AGRS.

Char

A total of 61 runs were made. The initial runs have been described by Ferrell et al.(1982) and McCue (1979). The last 38 runs of the series, designated by run numbers GO-26 through GO-61, provided a research data base. Runs GO-44A, 50A, and 50B were aborted due to failures of the coal feed screw. Pluggage of the venturi scrubber forced an early shutdown of run GO-52A, while difficulties with the fluidized bed temperature control caused the abortion of run GO-52B. The other runs were successfully completed.

Coal/Char Mixtures

Two runs, GO-32 and GO-33, were carried out with New Mexico coal but failed due to plugged exit lines. Based on this experience, the PCS system was modified to accommodate the tars associated with raw coal feedstocks. Details of the modification have been given by Ganesan (1980), and are summarized below.

- o movement of the venturi scrubber from the cyclone exit to the inlet of the PCS tank

- o enlargement of the line from the cyclone to the venturi from 1 inch to 3 inch pipe

- o installation of three quench spray nozzles to replace the original single nozzle
- o removal of the heat exchanger between the cyclone and the PCS tank and installation of cooling coils in the PCS tank
- o installation of an additional water separator downstream of the PCS tank to compensate for increased water entrainment of the modified system

Five runs using mixtures of the Western Kentucky char and New Mexico subbituminous coal were performed to test the ability of the modified facility to handle the high tars associated with a raw coal feed. The modifications were tested during runs GO-62 through GO-65. These runs used increasing fractions of New Mexico coal (from 10% to 50%), with the balance being Western Kentucky char. Four of the five attempted runs were completed. Run GO-63A was aborted due to failure of the 10" bed thermocouple. The completed runs indicated that the new PCS system could handle raw coal feedstocks, thus the run set was terminated and a set of runs with 100% coal was commenced.

New Mexico Coal

The New Mexico coal was much more difficult to gasify than were the char and coal/char mixtures. A run completion rate of 60% (18 out of 30) was achieved. The main operating problems were tar condensation during startup and feed screw stoppages. During startup the system was cold, and the tars formed condensed in the reactor exit lines and the cyclone. In several runs, the condensation caused severe pluggage and pressure buildup in the reactor. In most cases when this occurred, the pluggage was eliminated when the temperature reached approximately 650°F. The screw feeder stopped during several runs. Runs GO-67A, 69A, 75, 80A, and 81A were aborted when the screw could not be restarted. Removal screw failures forced the shutdown of runs GO-71A (packing blowout) and 85A (broken pin). A combination of high temperatures (in excess of 2000°F) causing clinkers, high pressures due to tar condensation, and gas leaks forced the abortion of runs GO-67B, 68A, 73A, and 77.

North Carolina Peat

The first two runs, GOP-1 and GOP-2, were carried out primarily for the purpose of finding a reasonable range of operating conditions for subsequent runs. Five of the eight attempted peat runs were completed. The failed runs were as follows: GOP-4A, due to failure of the packing

on the removal screw; GOP-5A, due to pluggage of the feed screw channel; and GOP-5B, due to failure of the feed and removal hydraulic systems.

The peat was a difficult material to feed with a screw conveyor, with stoppage of the screw occurring during several runs. The suspected cause was flow of wet gases into the screw channel, resulting in a wet paste-like feed that the screw feeder system could not handle. In several instances, the problem was cured by stopping the screw, increasing the screw purge nitrogen rate, and allowing the feed material to dry before restarting the screw. In addition, achieving a 38 inch bed was difficult in all runs due to the low ash content and high reactivity of the peat. The actual bed level was therefore estimated from bed pressure drop and temperature profile data.

Texas Lignite

Fourteen runs were made with Texas lignite as the feedstock. As for peat and coal, the first two attempted runs were exploratory runs to determine the feasible range of operating conditions for this feedstock. The following five runs, GOL-2 through 6, were successful runs, suggesting that the lignite might be an easy material to gasify in the NC State gasifier. However, four of the next six runs failed -- GOL-7A, 9A, and 10 due to clinker formation, and GOL-9B due to a combination of a feed screw stoppage and high bed temperatures. For GOL-9B the initial cause of the upsets was pressure fluctuations resulting from an operator's error. Run GOL-8, though completed, never achieved a well-behaved fluidized bed, and a steam supply failure forced an early shutdown. Due to these problems, the run was not included in data correlations and modeling efforts.

Following the failure of run GOL-10, the reactor startup procedure was reviewed and modified slightly. The initial reactor charge was changed from 15 lb of the Kentucky char to 10 lb of char plus 15 lb of a 50/50 mixture of char and lignite. The increase was designed to accommodate higher gas velocities during startup without jet penetration through the bed. The nitrogen flow rate during startup was increased to promote better mixing and avoid hot-spots. Using this modified procedure, run GOL-11 was successful. The completion of GOL-11 resulted in a total of seven successful lignite runs to be considered in further data analysis.

MASS BALANCES

Mass balances were performed following each run to assist in data evaluation. The attainment of steady-state operation was determined by inspection of gas analyses and plots of key variables versus time. A computer program was then used to average the data stored by the data

acquisition system over the specified steady-state period. The average temperatures, pressures, and flow rates were then used in a mass balance/data reduction program. The program used has beem described by Ferrell, et al. (1982). The output contains the following main components:

- o Reactor operating conditions - average bed temperature, pressure, apparent bed density and void fraction, solids holdup, and bed expansion factor.
- o Coal feed properties - coal type, proximate and ultimate analyses, average particle diameter, and particle and settled bed densities.
- o Feed rates of coal, steam, oxygen, and nitrogen, feed ratios, inlet temperatures, gas velocity, and average gas and solid residence times.
- o Output rates and compositions - make gas, spent char, cyclone dust, and tar.
- o Carbon and sulfur conversions.
- o Material balances on total mass, and the elements C, H, O, N, and S.
- o Results of wastewater and trace element analyses, and trace element material balances.

Program outputs for all runs are given by Purdy (1982). In general, the mass balance results were excellent, indicating that there were no gross errors in process variable or chemical analysis data. Following inspection of the mass balance results, it was arbitrarily decided that for a run to be acceptable total mass recoveries must be within 8% of 100%, carbon and oxygen recoveries within 10%, and hydrogen recoveries within 15%. Nitrogen recoveries were quite erratic due to known purge nitrogen leakage. In some runs, the purge rate was adjusted to close the nitrogen balance. Sulfur balances were also difficult to close completely due to the low amounts of sulfur. Table 8 lists the calculated total and elemental mass recoveries for all runs. Those runs which did not have acceptable mass balance results are indicated in Table 8, and have been omitted from the data correlations and modeling results to follow.

TABLE 8
MASS BALANCE RESULTS

Run	Total Mass	% Recovery C	H	O	N	S
KY Char						
GO-26	99.7	103.0	104.8	99.5	100.0	81.2
GO-27	103.7	107.2	106.5	103.9	100.0	99.2
GO-28	103.5	97.0	106.2	107.7	100.0	90.9
*GO-29	94.2	87.9	98.5	94.3	100.0	85.9
GO-30	99.3	93.3	102.9	100.0	100.0	103.9
GO-31	102.4	94.2	106.7	102.7	99.9	103.0
GO-34	97.6	96.8	107.4	95.7	100.0	115.5
GO-35	100.0	95.4	106.8	103.4	100.0	108.2
*GO-36	98.8	106.5	118.0	90.5	100.0	103.0
GO-37	97.6	100.4	109.1	94.1	100.0	98.3
GO-38A	95.5	99.1	97.5	94.3	100.0	89.3
GO-38B	95.6	96.4	100.5	95.6	100.0	86.9
*GO-39	90.1	82.0	99.5	84.5	100.0	111.2
*GO-40	79.5	89.2	72.6	65.8	100.0	98.7
*GO-41	95.5	106.2	98.4	88.6	100.0	114.2
GO-42	95.3	94.9	106.8	90.4	100.0	120.6
GO-43	103.7	106.6	92.2	106.6	100.0	94.1
GO-44B	98.9	99.5	101.8	99.3	100.0	94.3
GO-45	99.1	101.6	103.1	97.0	100.0	98.7
GO-46	96.3	90.0	102.2	96.3	100.0	97.1
GO-47	97.0	97.4	95.6	95.6	100.0	94.4
GO-48	102.2	104.6	104.6	103.1	100.0	99.4
GO-49	103.2	100.3	88.6	101.6	100.0	91.2
GO-51	104.1	99.5	108.4	108.5	100.1	98.8
*GO-53	101.3	92.2	119.5	105.7	100.1	96.8
GO-54	98.6	102.0	95.9	95.7	100.0	101.2
GO-55	101.8	106.0	101.2	102.4	100.0	109.5
GO-56	98.2	99.2	97.9	98.3	100.0	112.8
GO-57	98.8	100.7	92.5	97.7	100.0	107.0
GO-58	102.8	107.6	99.8	102.3	100.0	90.3
GO-59	98.4	99.9	97.6	98.2	100.0	103.8
GO-60	101.7	93.7	103.5	99.1	100.0	121.0
GO-61	96.4	92.5	99.3	95.2	100.0	114.8
Char/Coal Mixtures						
*GO-62	102.6	88.7	113.3	109.9	100.0	104.1
GO-63B	99.6	97.4	97.3	102.4	100.0	94.6
*GO-64	103.4	99.2	98.8	111.0	100.0	103.9
GO-65	99.7	105.7	93.8	101.0	100.0	108.5

* unacceptable mass balance

TABLE 8 (CONT.)

Run	Total Mass	C	H	O	N	S
NM Coal						
GO-66	95.5	107.3	100.4	99.9	85.4	118.8
GO-68B	102.5	108.5	101.4	99.8	108.8	107.7
GO-69B	97.7	105.0	106.8	97.9	110.7	95.1
*GO-70	100.5	117.6	104.1	93.6	115.8	101.4
*GO-71B	106.4	115.8	110.6	100.1	99.8	158.8
GO-72	101.2	110.0	107.2	102.2	104.7	129.1
*GO-73B	103.6	112.9	107.7	104.3	106.1	70.7
GO-74	98.0	101.6	96.6	97.7	106.5	87.8
GO-76	98.6	109.2	109.1	99.8	96.4	87.8
GO-78	102.0	109.1	98.6	98.8	99.0	102.1
GO-79	102.4	102.4	104.6	105.5	103.7	89.6
GO-80B	92.2	107.9	107.9	99.3	100.7	100.6
GO-81B	93.7	100.9	96.7	92.6	97.2	98.3
GO-81D	98.8	102.7	104.1	100.8	97.6	107.0
GO-82	96.1	105.4	99.9	99.3	98.3	112.2
*GO-83	103.9	82.1	98.4	107.2	99.7	96.4
GO-84	100.0	101.8	101.9	101.5	100.1	115.0
GO-85B	96.8	98.6	100.9	106.8	100.0	84.1
NC Peat						
GOP-1	99.2	100.7	108.3	96.0	100.3	84.4
*GOP-2	94.3	80.0	130.3	97.2	99.1	438.8
GOP-3	98.1	94.5	111.4	99.2	100.3	107.0
GOP-4B	100.7	90.3	109.4	100.4	100.1	86.8
GOP-6	98.3	97.3	105.6	95.1	99.8	104.9
TX Lignite						
GOL-2	99.4	102.8	101.9	101.9	105.6	104.8
GOL-3	100.7	105.0	111.6	108.1	99.0	95.7
GOL-4	96.2	98.7	112.6	96.7	102.7	108.2
GOL-5	96.9	106.7	108.0	101.6	91.1	99.5
GOL-6	98.5	100.0	109.4	98.0	117.3	88.7
GOL-7B	100.0	93.2	105.8	101.3	114.8	85.2
GOL-11	103.0	94.5	111.6	100.1	102.9	66.7

* unacceptable mass balance

CARBON CONVERSION AND GAS PRODUCTION

Carbon conversion is an important factor in any coal conversion process. For economical operation, it is crucial to achieve a high conversion. The carbon conversions reported in this study were calculated as the percent of the carbon in the feed coal that was converted to a gaseous form. The carbon contents of the tar and cyclone dust were not included in the calculated conversions. Figure 5 plots carbon conversion versus temperature for all gasifier runs. Due to varying run conditions there is a lot of scatter. However, the plot is not intended to show the effect of temperature, but instead to show the range of conversions achieved and the differences in the four gasifier feed materials. The plot suggests a ranking of the relative reactivities of the materials, lignite being the highest, followed by coal, peat, and char. The suggested relative reactivity for NC peat is deceptively low, as Figure 5 fails to account for the much lower solids residence times of the peat runs.

The reactor gas production rate was measured by an orifice meter at the exit of the PCS system. The measured make gas production rate varied from over 20 scfm for the raw coal feedstocks to less than 10 scfm for the KY char. Examination of the gas production rate data gives the same qualitative information as Figure 5 -- low relative reactivity for the char, and high relative reactivities for the other feeds.

The effect of reactor operating parameters on carbon conversion and the gas production rate was difficult to determine from the experimental results. The difficulty resulted from the interdependence of the operating parameters -- for example, varying the reactor temperature required changing the oxygen feed rate. In a similar manner, variation of any operating parameter resulted in the variation of one or more other operating parameters. However, it was possible to select sets of runs for which the operating parameters other than that of interest showed only small variances. In this manner the approximate effects of reactor operating parameters were examined.

Effect of Temperature

The effect of temperature on carbon conversion for Kentucky char is shown in Figure 6. The molar H_2O/C feed ratio (0.7-0.9), char feed rate (33-44 lb/hr), and bed height (38 in) were not varied greatly for the runs included. The runs at 1680-1700^{o}F had conversions of only 16-22%, while the runs at 1780-1820^{o}F had conversions of 31-47%, roughly double that of the lower temperature runs. The gas production rate also increased with increasing temperature.

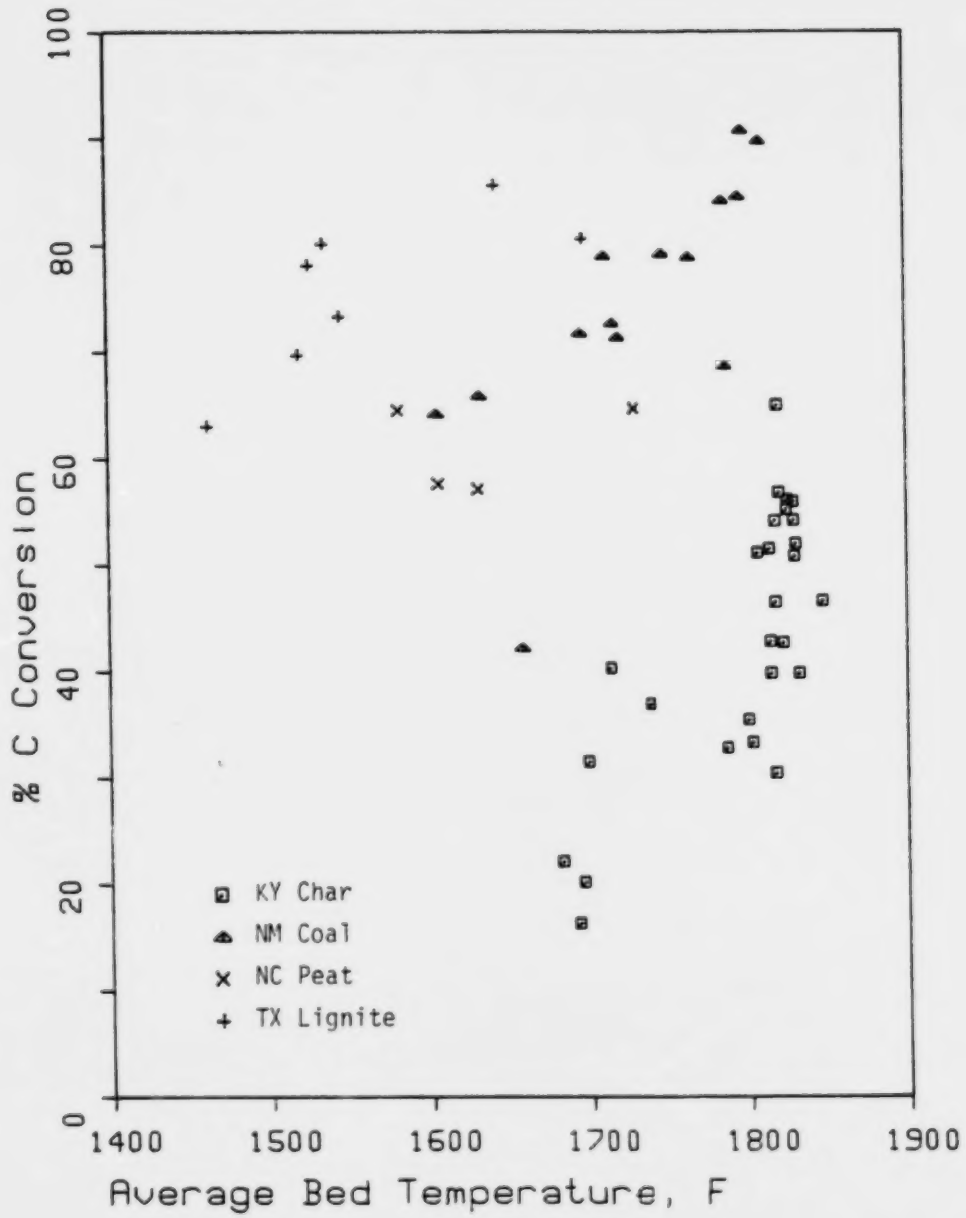

FIGURE 5

CARBON CONVERSION VERSUS TEMPERATURE

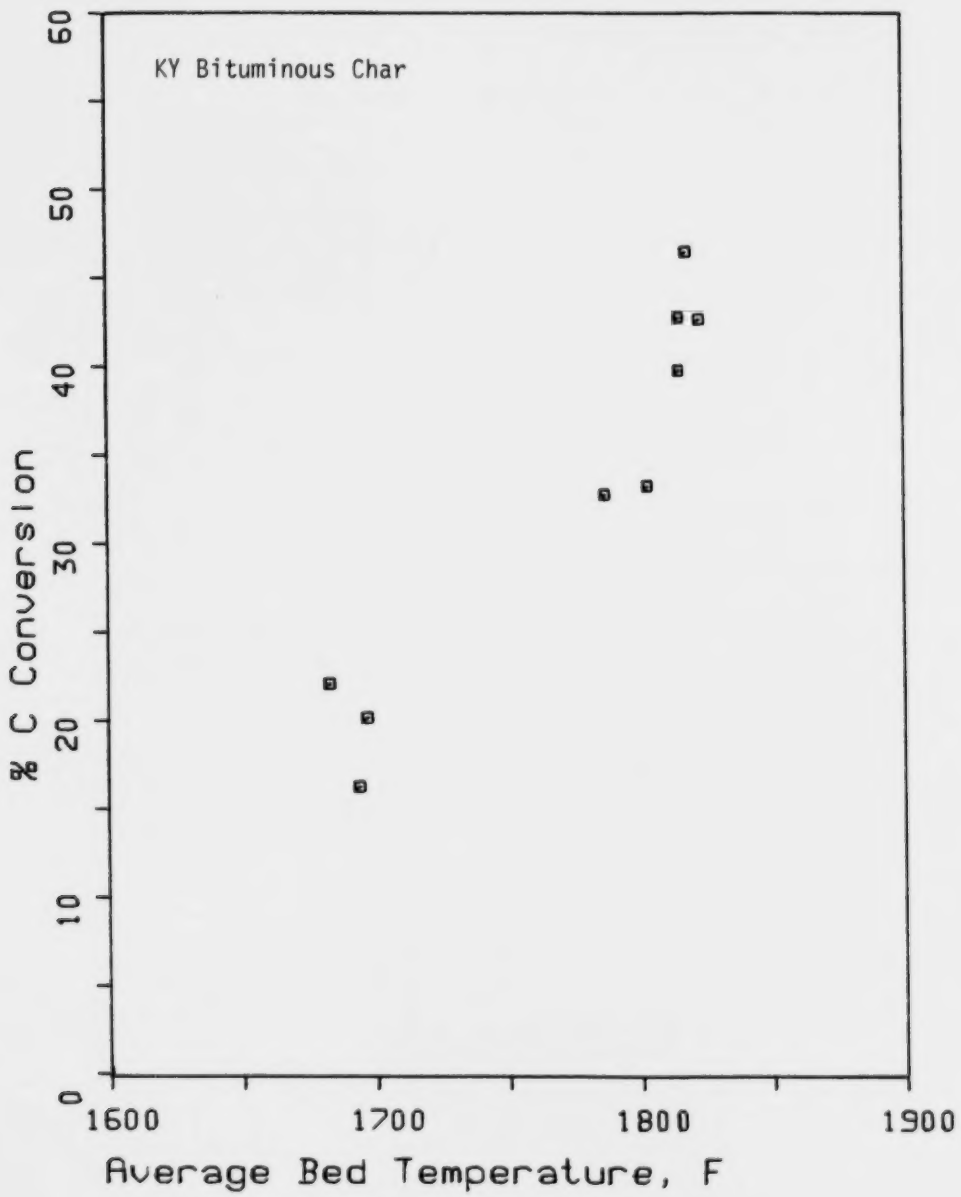

FIGURE 6

EFFECT OF AVERAGE BED TEMPERATURE ON CARBON CONVERSION FOR KENTUCKY CHAR

Figure 7 plots carbon conversion versus average bed temperature for a set of New Mexico coal runs. The data is from runs having similar pressures (101-107 psig), coal feed rates (41-49 lb/hr), steam feed rates, and bed heights (30-38 in). The dry N_2-free gas production rate for the same set of runs is plotted versus the average bed temperature in Figure 8. The plots show that the carbon conversion and the dry N_2-free gas rate both increase with increasing temperature.

Increasing temperature increases the gasification reaction rates, thus giving the observed increases in conversion and gas rates. For both the char and the coal, increased temperatures were obtained by increasing the oxygen feed rate. Thus, a part of the observed increases in carbon conversion and gas production were due to greater carbon combustion to yield carbon monoxide and carbon dioxide.

Effect of Steam/Carbon Feed Ratio

A set of char runs with similar temperatures (1800-1830^{o}F), char feed rates (33-38 lb/hr), pressure (101-106 psig), and bed height (38 in) was chosen to demonstrate the effect of the molar H_2O/C feed ratio on carbon conversion and gas production. Figure 9 shows the effect on carbon conversion, and Figure 10 shows the effect on the dry N_2-free gas production rate. Both show increases with increasing H_2O/C ratios.

Increasing the H_2O/C ratio enhances the carbon-steam reaction, thus giving greater conversion and gas production. Increased steam also increases the driving force force for the water gas shift reaction toward the dry gases CO_2 and H_2. In addition, the greater amount of steam requires slight increases in oxygen for temperature maintenance, as indicated in Figure 11. Thus, greater conversion and gas production may occur by combustion.

For the New Mexico coal insufficient runs were available to isolate the effect of the H_2O/C feed ratio.

Effect of Pressure

The effect of pressure cannot be determined for the New Mexico coal and the North Carolina peat as all runs were made at roughly 100 psig. However, three low pressure runs were made with the Kentucky char. Two of the low pressure runs, GO-49 and GO-51, had operating conditions other than pressure nearly identical to the high pressure runs GO-46 and GO-47. The gas velocities were held constant at about 1 ft/s in each run by adjustment of the feed nitrogen rate. Table 9 compares the conditions and primary results for the four runs. Carbon conversion and gas production were nearly identical for all of the runs. In addition, all other output variables for the high and low pressure runs showed good agreement.

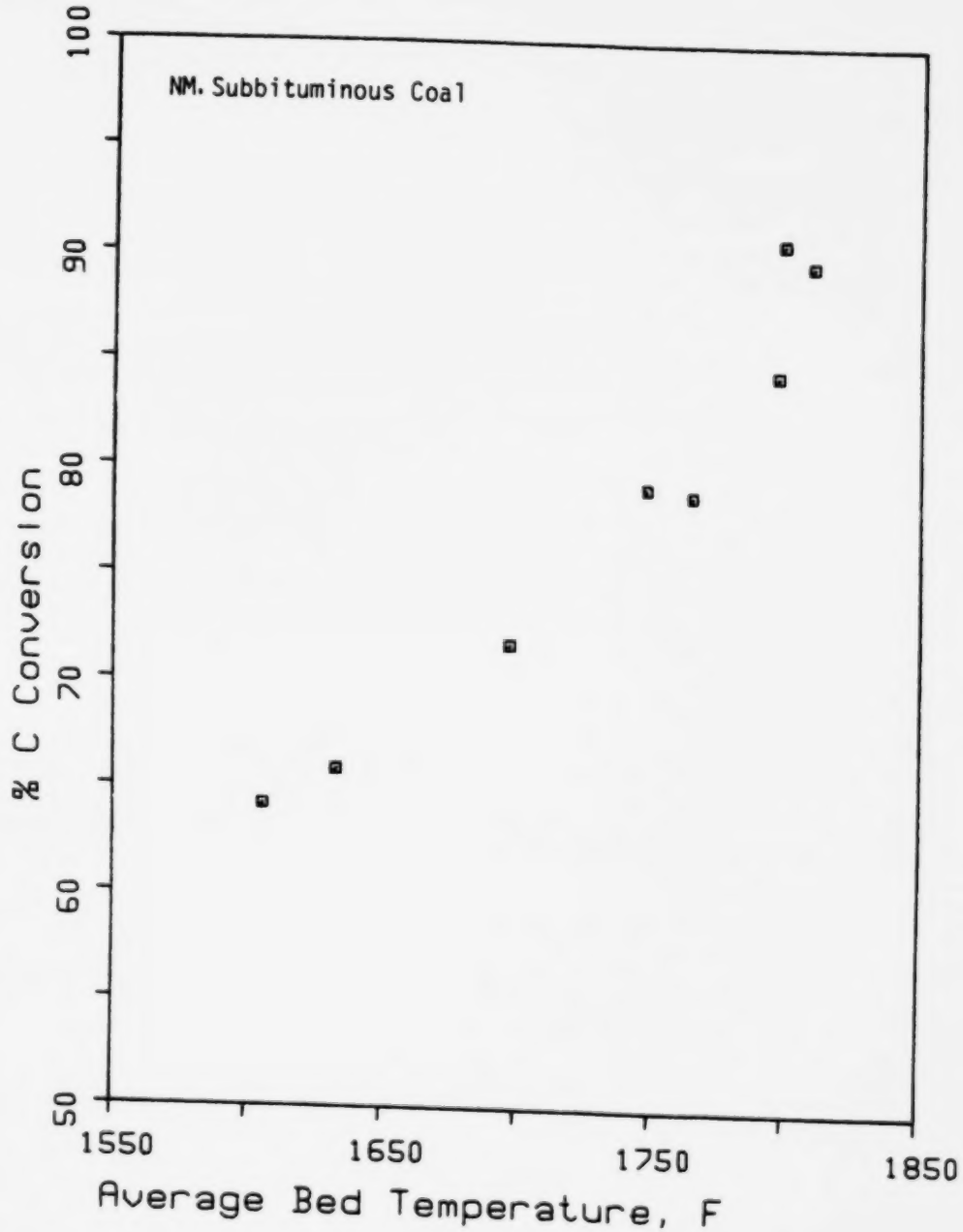

FIGURE 7

EFFECT OF AVERAGE BED TEMPERATURE ON CARBON CONVERSION FOR NEW MEXICO COAL

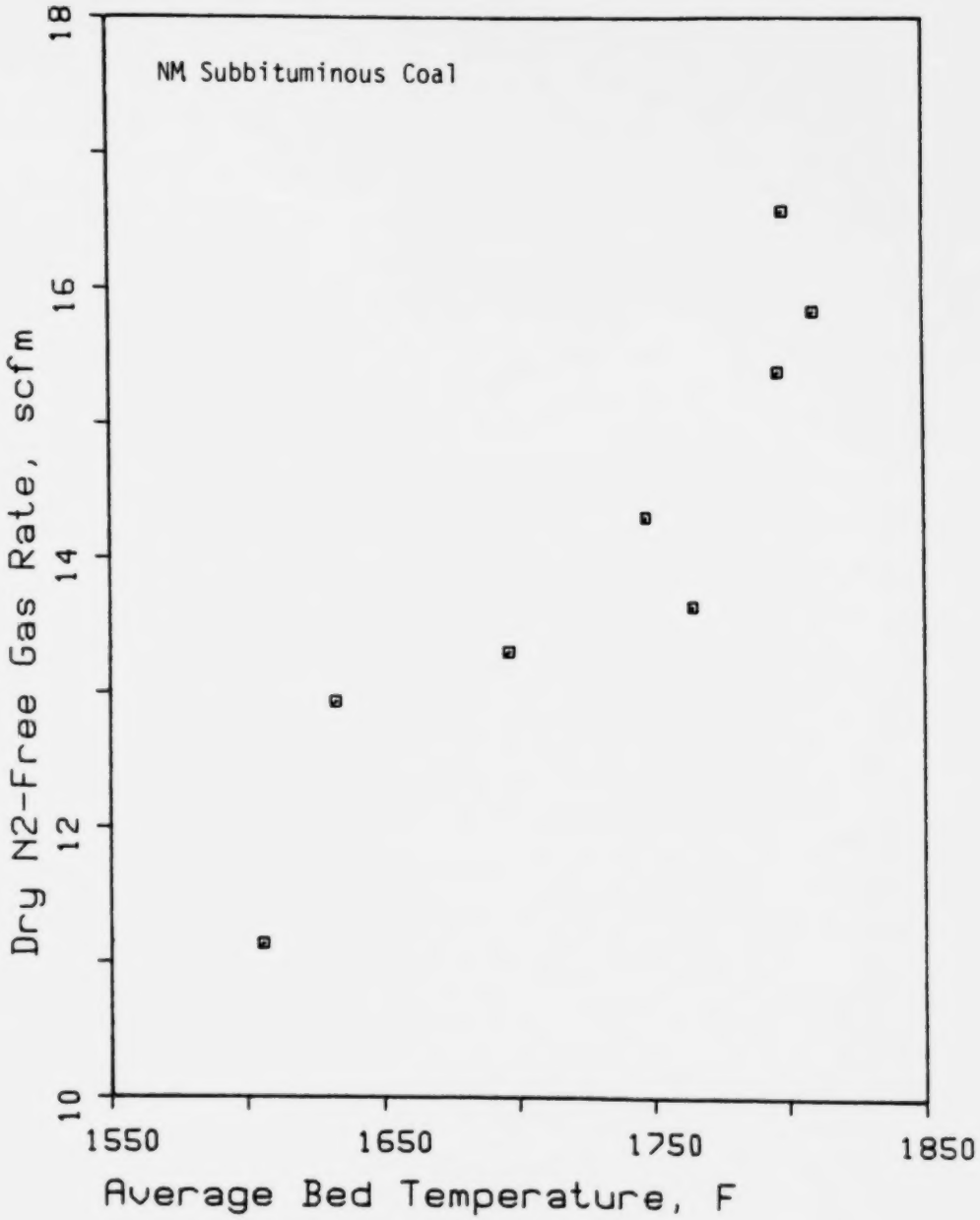

FIGURE 8

EFFECT OF AVERAGE BED TEMPERATURE ON GAS PRODUCTION

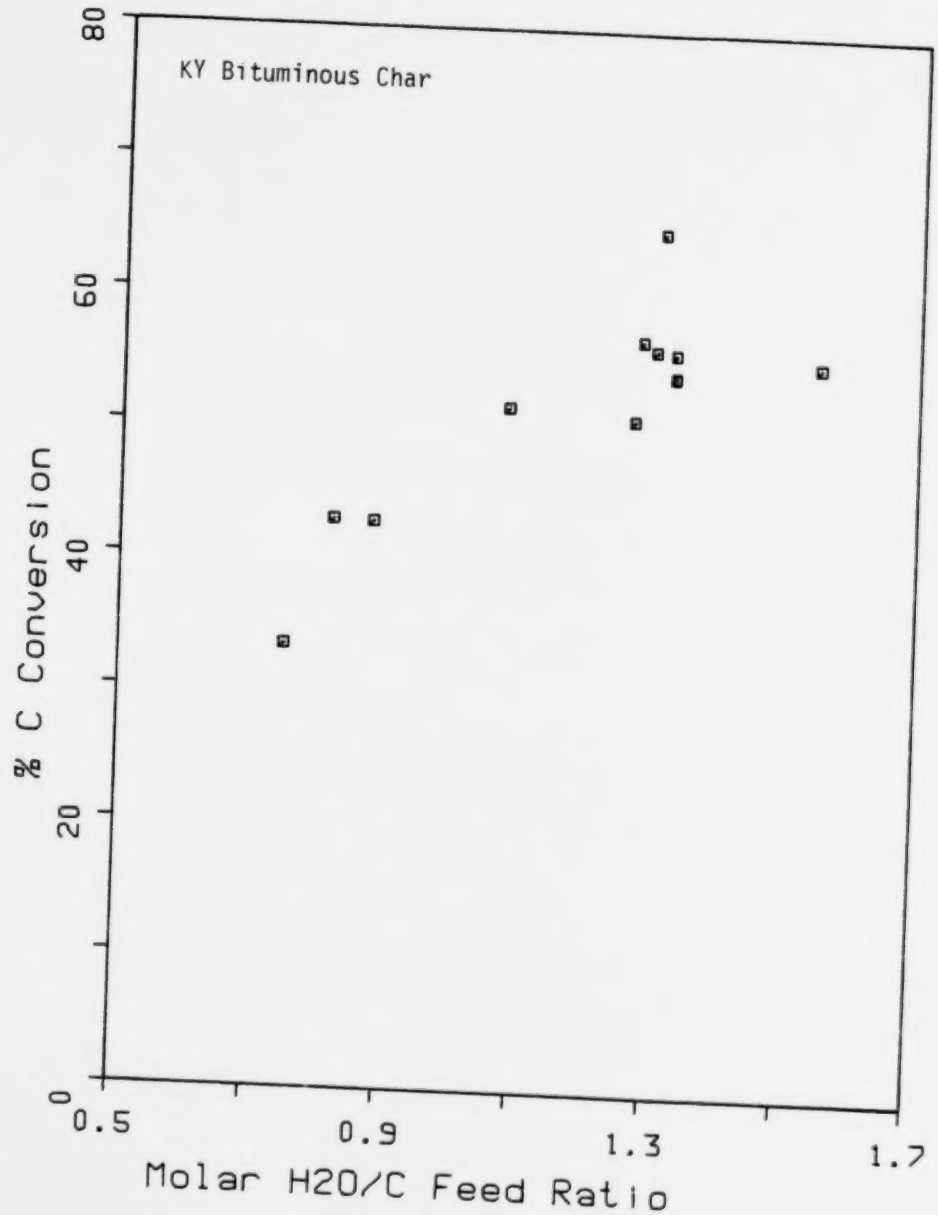

FIGURE 9

EFFECT OF STEAM-TO-CARBON FEED RATIO
ON CARBON CONVERSION

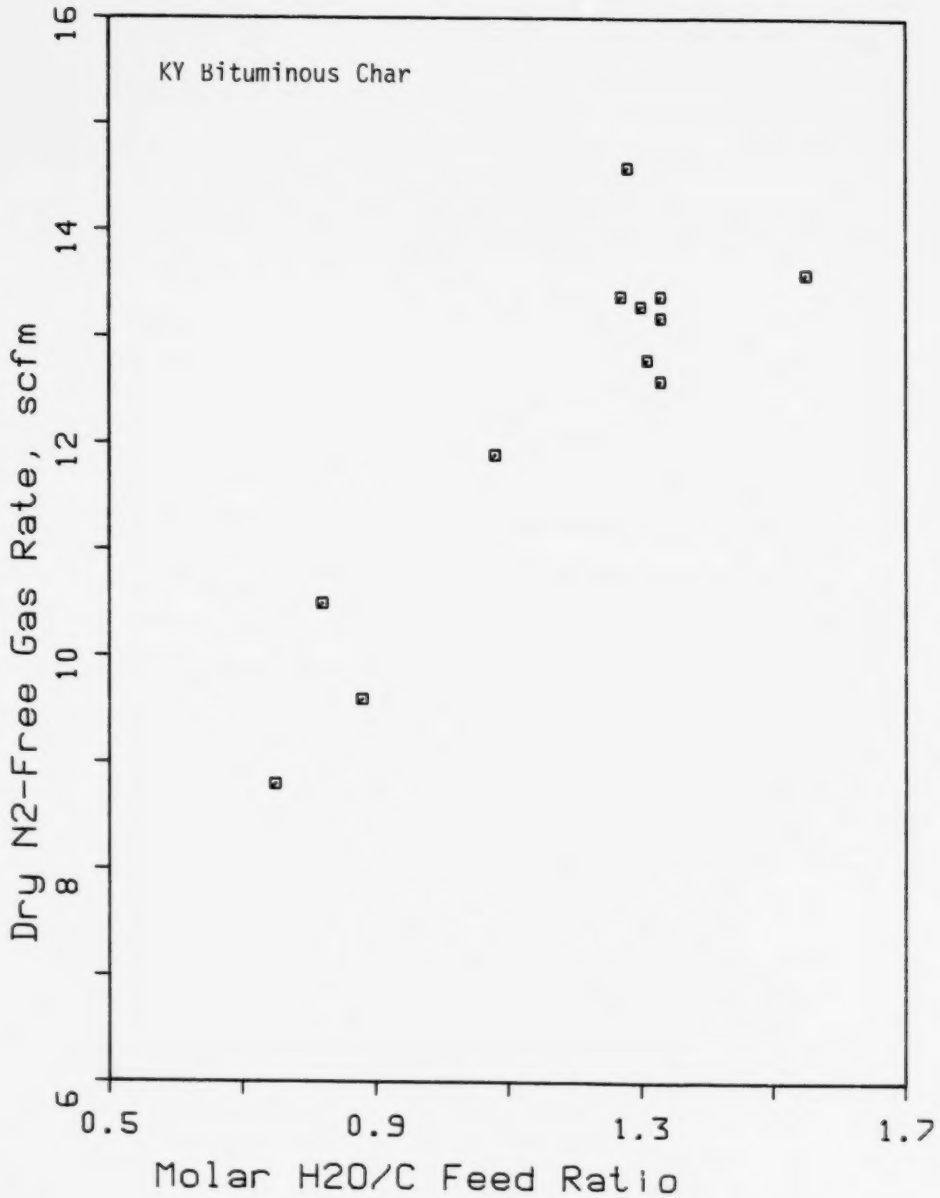

FIGURE 10

EFFECT OF STEAM-TO-CARBON FEED RATIO
ON GAS PRODUCTION

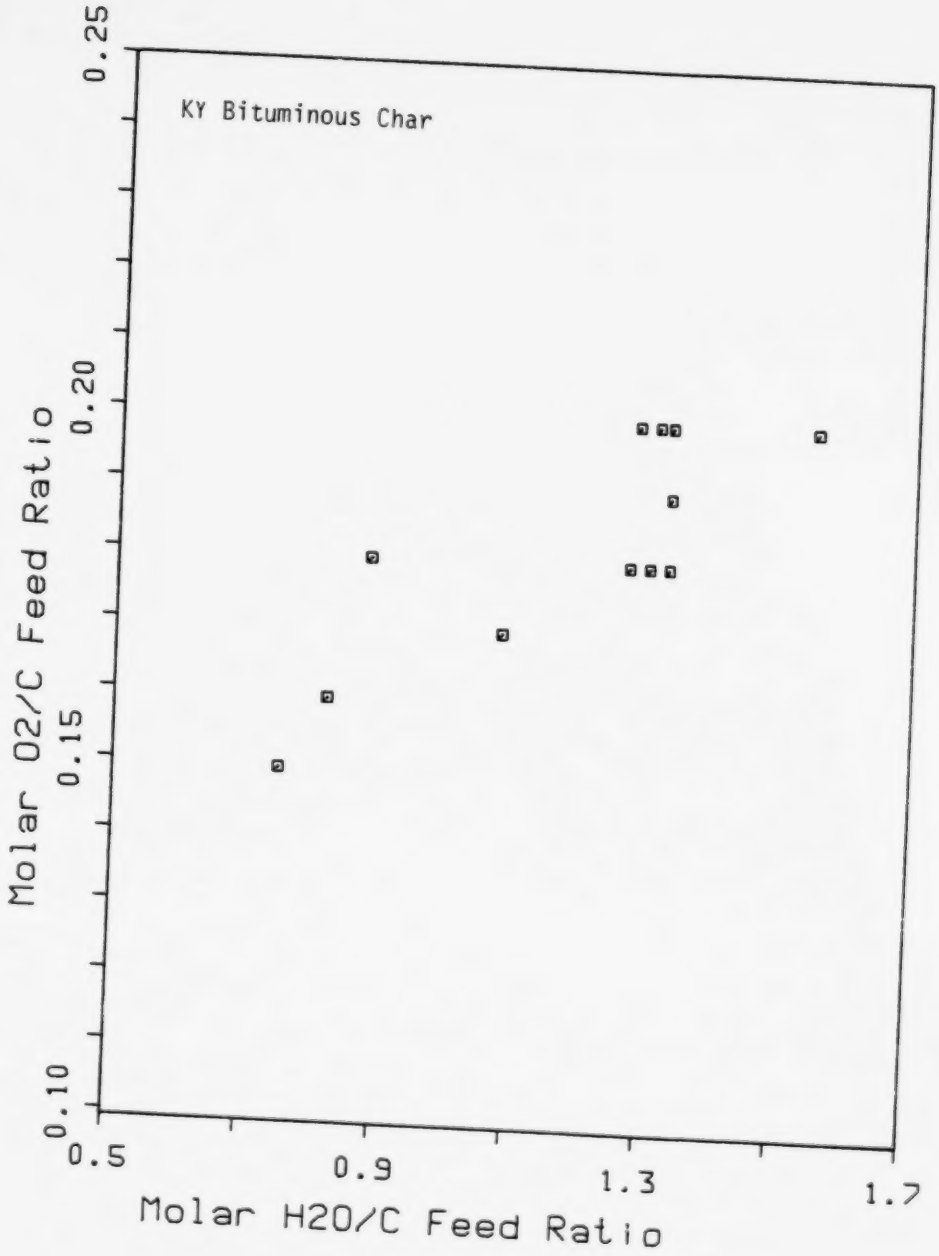

FIGURE 11

EFFECT OF STEAM-TO-CARBON FEED RATIO
ON OXYGEN-TO-CARBON FEED RATIO

TABLE 9
EFFECT OF PRESSURE ON CONVERSION AND GAS PRODUCTION

Run	Pressure psig	Temp °F	H_2O/C	Char Feed lb/hr	C Conv %	Dry N_2-free Gas, scfm
GO-46	104	1823	0.88	33.2	42.8	9.6
GO-47	101	1815	0.82	36.1	42.9	10.5
GO-49	78	1818	0.83	36.7	46.6	10.0
GO-51	78	1815	0.83	37.0	39.9	10.4

Effect of Solids Residence Time

The average solids residence time was calculated by determining the total mass of the bed from pressure drop measurements across the fluidized bed. The residence times calculated in this manner ranged from 15 to 47 min for the Kentucky char, 8-22 min for the New Mexico coal, 0.8-2.4 min for the peat, and 7-13 min for the lignite. The low values for peat were a result of the peat's high reactivity and low ash content.

The average solids residence time in the reactor depends primarily on the fluidized bed height and the coal feed rate. Variations in the bed porosity can also influence the residence time; however, the porosity varied little and thus was not a major influence on the solids residence time.

The bed height was not varied significantly for the coal, peat, or lignite runs, and determining the bed height effect is difficult. However, for the Kentucky char, runs GO-26 and GO-44B can be compared to see the effect of bed height. Run GO-26 had a bed height of 52 in, while run GO-44B had a bed height of 38 in. The other operating conditions were roughly the same for the two runs. Table 10 summarizes the principle conditions and results for the two runs. As would be expected, the greater bed height, and thus a greater residence time, resulted in a greater carbon conversion and gas production rate.

TABLE 10
EFFECT OF BED HEIGHT ON CONVERSION AND GAS PRODUCTION

Run	Res Time, min	Temp °F	H_2O/C	Char Feed lb/hr	C Conv %	Dry N_2-free Gas, scfm
GO-26	47	1714	1.0	30.8	40.4	8.2
GO-44B	33	1700	1.3	34.7	31.6	7.4

The effect of the coal feed rate can be seen by comparing char runs GO-46 and GO-54. For these two runs, the bed heights were the same at 38 in, and the only major difference was the feed rate. Run GO-54 had nearly double the feed rate of GO-46, and therefore nearly one-half the average solids residence time of run GO-46. The result is a lower conversion, but a higher gas production rate for GO-54. The lower residence time gives the lower carbon conversion, and the higher throughput rate and lower conversion gives a more reactive material in the bed, resulting in the greater gas rate. The run conditions, conversions, and gas rates are given for both runs in Table 11.

TABLE 11
EFFECT OF COAL FEED RATE ON CONVERSION AND GAS PRODUCTION

Run	Res Time, min	Temp °F	H_2O/C	Char Feed lb/hr	C Conv %	Dry N_2-free Gas, scfm
GO-46	30	1823	0.88	33.2	42.8	9.6
GO-54	16	1801	0.87	64.9	35.5	15.3

REACTOR EFFLUENTS

Gases

The gas emerging from the gasifier contained a large amount of unconverted steam which was condensed by the PCS system. The mole percent of water ranged from 30% to 60% for char gasification, 35% to 53% for the New Mexico coal, 40% to 50% for peat, and 35% to 55% for the lignite runs. Following the quenching by the PCS system, the gas was filtered, dehydrated, and compressed before entering the acid gas removal system.

Table 12 gives average dry nitrogen-free gas compositions measured for the gasification of char, coal, peat, and lignite. For each feedstock, the gas composition showed little variance from run to run. Comparing feedstocks, the major difference is the lack of sulfur gases and hydrocarbons other than H_2S, COS, and methane for the Kentucky char. A much larger number of species were identified in the product gases from the gasification of the raw coals. In addition, the char yielded a gas of greater carbon monoxide content compared to the other materials. Sulfur gas content was roughly in proportion to the sulfur content of the feed materials -- greatest for the char, least for the peat. Among the raw coals, hydrocarbon content was greatest for the peat, followed by the NM coal, and least for the lignite.

The compositions shown in Table 12 are comparable to those measured by Forney et al.(1974) for the Synthane process. For a Wyoming subbituminous coal and a North Dakota lignite similar to the coals used in this study, hydrogen sulfide concentrations of 0.175-0.248% and carbonyl sulfide concentrations of 0.0032-0.0065% were measured. BTX measurements yielded the following results: 0.04-0.17% for benzene, 0.006-0.017% toluene, and 0.003-0.007% for C_8 aromatics (predominantly xylenes and ethylbenzene). The Synthane gasifier is a top-feeding fluidized bed similar to the NCSU gasifier; the similarity in sulfur and BTX gas concentrations is thus expected. The major gaseous products (CO, CO_2, CH_4, and H_2) from both processes also show similar concentrations (Elliot, 1981).

The water gas shift reaction is an important reaction in gasification. It has often been assumed to be at equilibrium in gasification processes. The equilibrium constant for the reaction is given by

$$K_{wg} = \frac{[CO_2][H_2]}{[CO][H_2O]} \qquad (2)$$

Experimentally measured concentrations of CO, CO_2, H_2, and H_2O were substituted into equation (2) and plotted vs the temperature at the top of the fluidized bed in Figure 12. All but one of the experimental values of K_{wg} fall below the equilibrium line, suggesting that the reaction is not at equilibrium. Modeling studies to be discussed later will take a further look at the question of shift reaction equilibrium.

Sulfur Gases--

Examination of run data indicated that the sulfur conversion could be roughly estimated by assuming it to equal the carbon conversion. Sulfur conversion is plotted versus carbon conversion in Figure 13 for the coal, char, lignite, and peat. Only data from runs having sulfur balances within 20% of closure were included. Sulfur conversions were generally greater than carbon conversions for char runs, but for the coal, peat, and lignite runs the two conversions were nearly the same.

TABLE 12
COMPARISON OF GAS COMPOSITIONS
Mole %, dry N2-free basis

	KY Char	NM Coal	NC Peat	TX Lignite
CO	26.0	21.4	21.5	18.8
H_2	40.9	40.0	37.3	41.6
CH_4	2.4	8.3	8.6	6.2
CO_2	29.7	28.3	30.2	32.1
H_2S	0.9700	0.3330	0.0670	0.2480
COS	0.0420	0.0100	0.0035	0.0060
CS_2	---	0.0003	0.0003	0.0001
CH_3SH	---	0.0025	0.0040	0.0035
C_2H_5SH	---	0.0003	---	0.0001
Thiophene	---	0.0030	0.0020	0.0025
Ethylene	---	0.5260	0.7200	0.3240
Ethane	---	0.6340	0.8960	0.4760
Propylene	---	0.1760	0.3880	0.1140
Propane	---	0.0530	0.1120	0.0460
Butylene	---	0.0620	---	0.0500
Butane	---	0.0280	0.0500	0.0160
Benzene	---	0.1020	0.1180	0.0670
Toluene	---	0.0600	0.0790	0.0200
Ethylbenzene	---	0.0015	---	0.0020
p-Xylene	---	---	---	0.0008
m-Xylene	---	0.0040	---	0.0035
o-Xylene	---	0.0020	---	0.0020

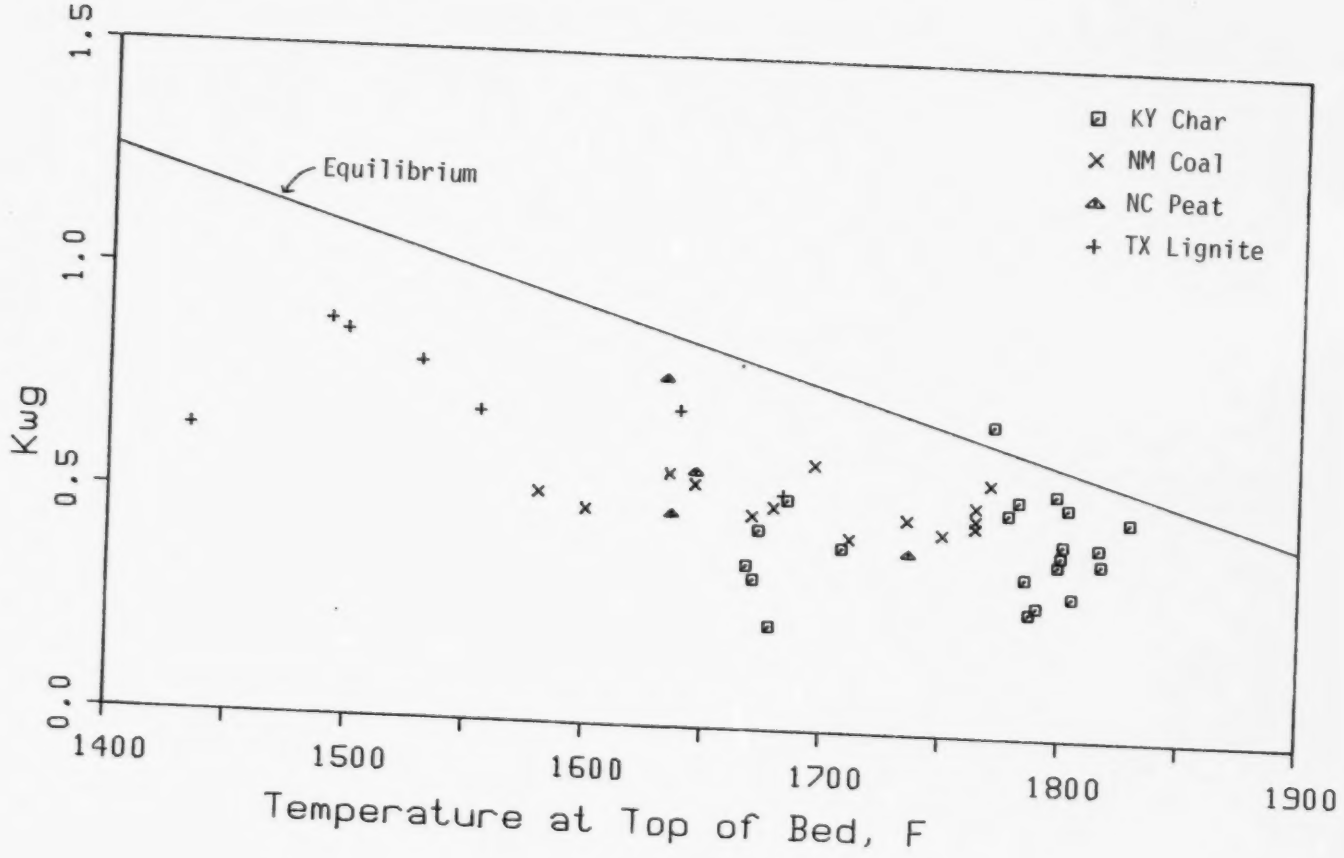

FIGURE 12

EXPERIMENTAL VALUES OF K_{WG} VERSUS TEMPERATURE

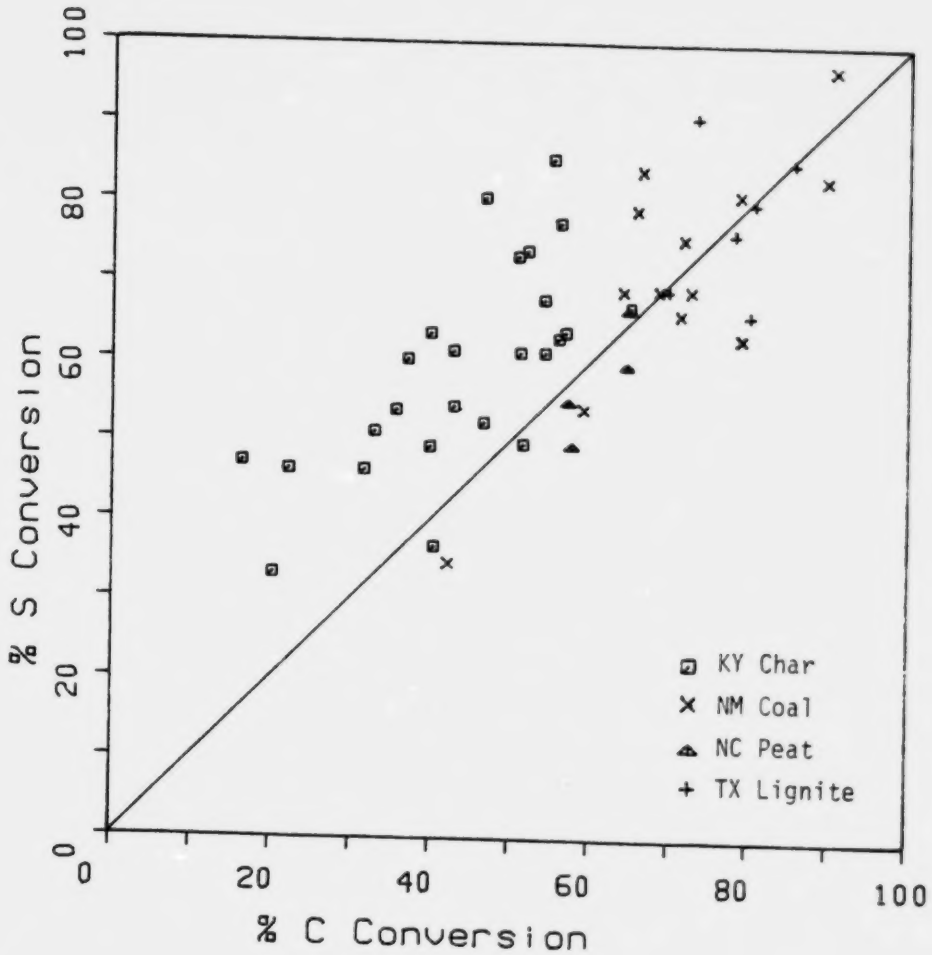

FIGURE 13
COMPARISON OF SULFUR AND CARBON CONVERSIONS

Sulfur gas production rates, calculated as grams emitted per kilogram of dry ash-free coal fed, are tabulated in Tables 13 and 14. Table 13 gives values for char gasification and Table 14 gives values for the gasification of coal, peat, and lignite. H_2S and COS were the only sulfur species detected during the gasification of the Western Kentucky char. The gasification of the raw coal feedstocks yielded carbon disulfide, mercaptans, and thiophene in addition to H_2S and COS. Sulfur dioxide was below the detection limit of approximately 20 ppm for all gasifier runs. The sulfur gas production levels were generally proportional to the dry ash-free sulfur contents of the feed coals: 3% for the char, 1.2% for the NM coal, 0.2% for the peat, and 1.0% for the lignite. However, the distributions of the sulfur species differ slightly. Table 15 presents average distributions of the gaseous sulfur for each feed. The major difference is the relatively high amounts of methyl mercaptan and thiophene for the peat.

TABLE 13
SULFUR GAS PRODUCTION RATES FROM CHAR GASIFICATION
Grams per kilogram of DAF feed

Run	Temp* (°F)	H_2S	COS	Run	Temp* (°F)	H_2S	COS
GO-26	1714	12.0	0.2	GO-45	1697	9.6	1.4
GO-27	1807	19.5	1.5	GO-46	1823	18.4	1.3
GO-28	1821	19.6	1.3	GO-47	1815	16.5	1.3
GO-30	1830	21.2	2.9	GO-48	1818	21.2	1.7
GO-31	1833	19.0	1.4	GO-49	1818	17.0	1.7
GO-34	1694	14.3	1.1	GO-51	1815	15.3	1.3
GO-35	1683	13.9	0.9	GO-54	1801	17.0	1.1
GO-37	1787	13.9	1.2	GO-55	1739	19.4	1.7
GO-38A	1830	22.3	1.5	GO-56	1825	27.5	2.2
GO-38B	1830	21.7	1.5	GO-57	1826	24.5	2.1
GO-42	1803	21.6	1.6	GO-58	1814	15.2	2.0
GO-43	1820	20.2	1.6	GO-59	1831	23.9	1.5
GO-44B	1700	14.1	1.1	GO-60	1818	19.4	1.3
				GO-61	1848	24.2	1.6

* volume-weighted average bed temperature

TABLE 14
SULFUR GAS PRODUCTION RATES FROM GASIFICATION
OF NM COAL, NC PEAT, AND TX LIGNITE
Grams per kilogram of DAF feed

Run	Temp* (°F)	H_2S	COS	CS_2	CH_3SH	C_2H_5SH	Thio
NM Coal							
GO-66	1799	10.7	0.72	0.016	0.029	---	---
GO-68B	1787	7.6	0.61	0.013	0.092	---	---
GO-69B	1720	9.3	0.43	0.007	0.123	---	0.22
GO-72	1797	12.6	0.52	0.026	0.070	0.015	---
GO-74	1659	4.1	0.21	0.005	0.042	---	0.11
GO-76	1712	7.6	0.41	0.020	0.077	---	0.17
GO-78	1606	7.6	0.41	0.007	0.231	---	0.23
GO-79	1748	7.5	0.54	0.016	0.186	0.007	0.88
GO-80B	1765	10.8	0.49	0.023	0.109	---	0.15
GO-81B	1717	8.4	0.50	0.015	0.081	0.18	0.29
GO-81D	1633	9.7	0.29	0.006	0.114	---	0.14
GO-82	1810	10.0	0.46	0.025	0.060	---	0.16
GO-84	1697	10.2	0.38	0.006	0.116	---	0.17
GO-85B	1786	8.4	0.47	0.024	0.130	---	0.15
NC Peat							
GOP-1	1730	1.36	0.131	0.013	0.167	---	0.154
GOP-3	1582	1.16	0.117	---	0.079	---	0.076
GOP-4B	1607	1.09	0.092	---	0.070	---	0.068
GOP-6	1632	1.46	0.098	---	0.106	---	0.083
TX Lignite							
GOL-2	1644	7.60	0.340	---	1.284	---	0.216
GOL-3	1699	8.05	0.517	---	0.200	0.014	0.239
GOL-4	1546	8.76	0.217	0.008	0.091	---	0.099
GOL-5	1527	7.30	0.558	---	0.172	---	0.125
GOL-6	1536	5.40	0.236	---	0.095	---	0.087
GOL-7B	1520	5.71	0.161	---	0.147	---	0.233
GOL-11	1463	4.36	0.168	---	0.119	---	0.114

* volume-weighted average bed temperature

TABLE 15
DISTRIBUTION OF GASIFIED SULFUR

S Form	Wt % of Total Gasified Sulfur			
	KY Char	NM Coal	NC Peat	TX Lignite
H_2S	95.8	95.3	87.1	95.2
COS	4.2	2.9	4.4	2.3
CS_2	--	0.14	0.70	0.09
CH_3SH	--	0.72	5.23	1.39
C_2H_5SH	--	0.07	--	0.05
Thiophene	--	0.90	2.61	0.93

Hydrogen sulfide production rates are plotted versus the average bed temperature in Figures 14 and 15. Despite scatter in the plots, it appears that the H_2S rate shows a slight increase with increasing temperature for char, but no apparent dependence for the other feeds. A substantial fraction of the total H_2S production rate occurs during devolatilization for the raw coals. This process takes place in the freeboard zone of the reactor where the temperature showed little variance from run to run, despite large variances in the fluidized bed temperature. For the devolatilized char, it is likely that all H_2S production occurred in the fluidized bed. This explains the apparent lack of dependence on temperature for the coal, peat, and lignite H_2S production rates. The other sulfur species production rates were much lower than the H_2S rates, and no trends with temperature could be observed.

It is likely that during steam-oxygen gasification the gas-solid reactions formed mainly hydrogen sulfide. Gas-phase reactions then tended to produce COS. The two gas-phase reactions of most importance involving H_2S and COS are:

1. $COS + H_2O = H_2S + CO_2$ (3)

2. $COS + H_2 = H_2S + CO$ (4)

The equilibrium constants for these two reactions are

$$K_1 = [H_2S][CO_2]/[COS][H_2O] \quad (5)$$

$$K_2 = [H_2S][CO]/[COS][H_2] \quad (6)$$

where the brackets indicate any convenient concentration units. Ideal gas behavior is assumed.

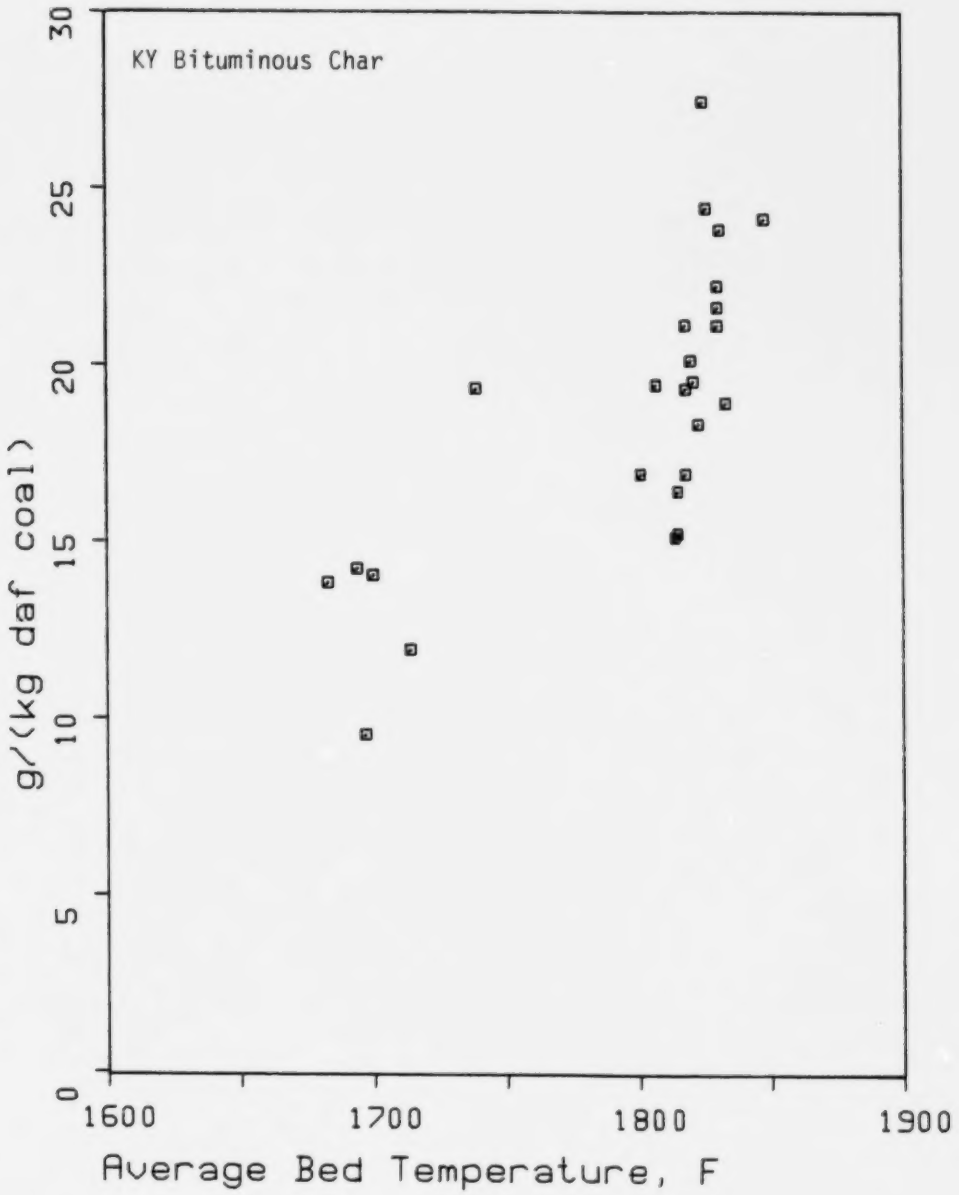

FIGURE 14

HYDROGEN SULFIDE PRODUCTION RATES FOR KY CHAR

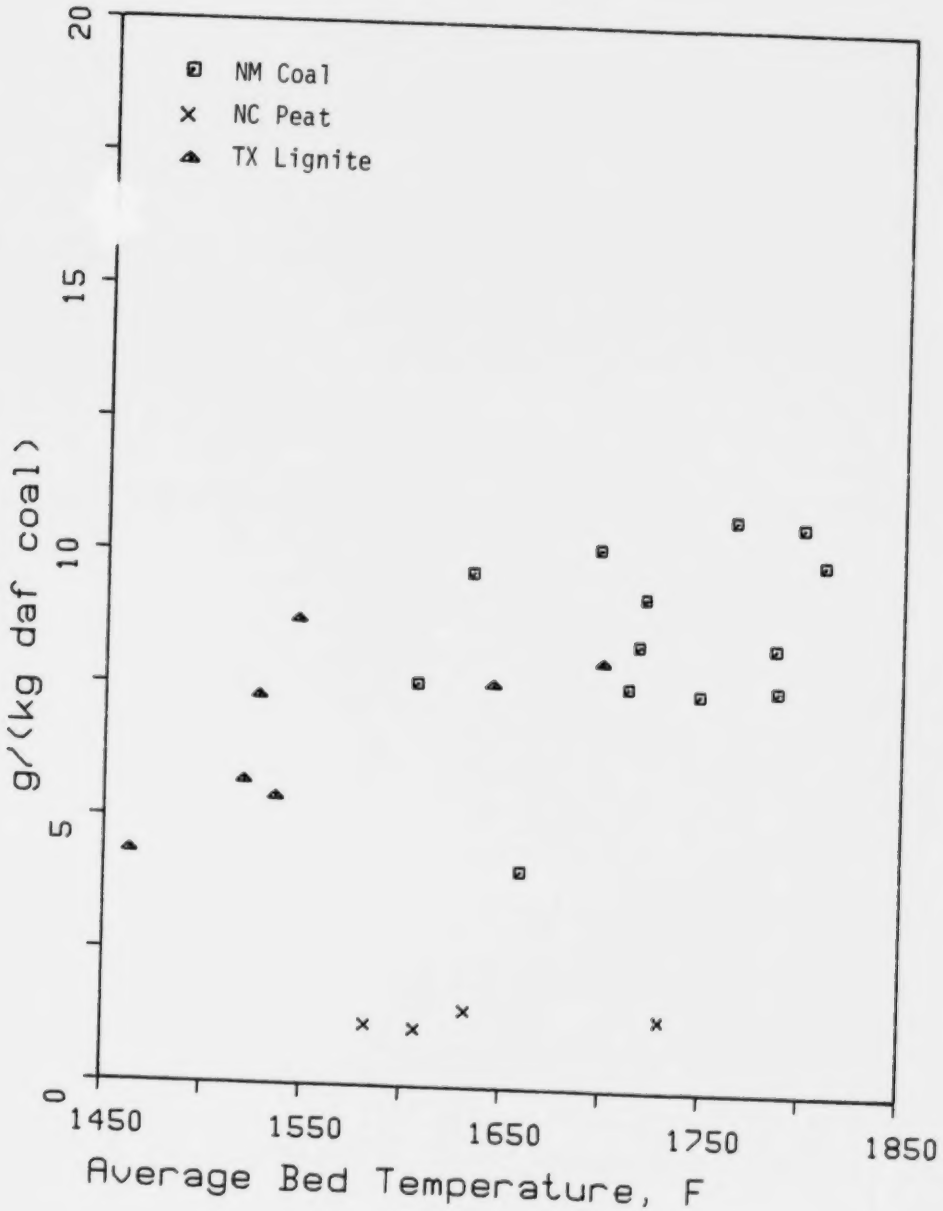

FIGURE 15

HYDROGEN SULFIDE PRODUCTION RATES FOR NM COAL, NC PEAT, AND TX LIGNITE

Figures 16 and 17 show plots of the experimental values of the constants K_1 and K_2 as defined above. Also shown on the graphs are lines representing the equilibrium data given by Kohl and Riesenfeld (1979). The plots indicate that both reactions are nearly at equilibrium. However, the sum of the two reactions is the water-gas shift reaction, for which previous data indicated equilbrium is not achieved. Therefore, reactions 1 and 2 cannot both be at equilibrium. The data for K_2 generally lies below the equilibrium curve; it is therefore likely that reaction 1 is at equilibrium while reaction 2 is not. In addition, the data from which these equilibrium lines were drawn were taken from two different sources, and there are likely inconsistencies between the two.

Gilman (1983) performed bench scale batch pyrolysis experiments as a part of the coal research program at NCSU. Figures 18-22 compare production rates of sulfur species from the gasification pilot plant with the pyrolysis rates measured by Gilman for New Mexico coal. The production rates of H_2S and COS were much higher for gasification than for pyrolyis. The difference may be attributable to the occurance of char sulfur-H_2 reactions in the gasifier. This is supported by the high H_2S and COS production rates measured for the gasification of the devolatilized Kentucky char. The other sulfur species -- thiophene, methyl mercaptan, and carbon disulfide -- showed similar production levels for both gasification and pyrolysis, indicating that they are produced primarily during devolatilization.

Hydrocarbon Gases--

In the gasification of the raw coal feedstocks a large number of hydrocarbon gases were produced. Aliphatics up to butene and butane and simple aromatic compounds were detected in the gasifier make gas stream. Table 16 summarizes the production rates of aliphatic hydrocarbons from coal, peat, and lignite gasification. Char gasification yielded no measurable hydrocarbons other than methane, for which production levels ranged from 10 to 30 grams per kilogram of DAF feed. Methane levels were much greater for the raw coals, with NM coal exhibiting the highest level. Hydrocarbons higher than methane showed the same general production levels for the coal, peat, and lignite.

Examination of the production rates as a function of temperature revealed few trends. Figure 23 plots the ethane and ethylene rates versus the average bed temperature for the Texas lignite. For this case, the rates show a clear increase with increasing temperature. However, in general there was much data scatter and little dependence on temperature could be noted. The lack of temperature dependence is most likely due to the same reason stated for sulfur gases -- most hydrocarbons of molecular weight higher than methane are evolved during devolatilization in the freeboard zone, where the temperature shows little variance from run to run.

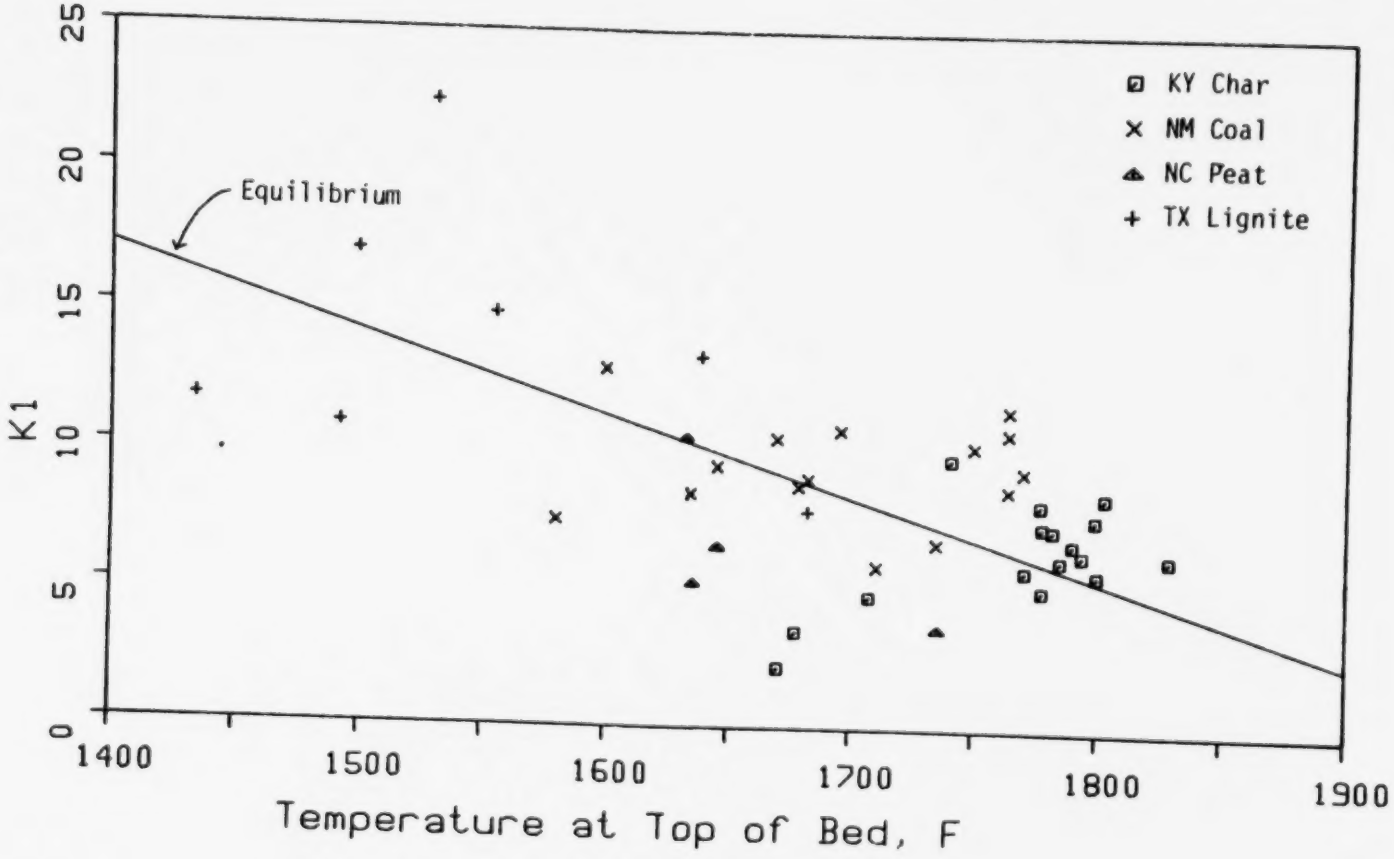

FIGURE T6

EXPERIMENTAL VALUES OF K_1 VERSUS TEMPERATURE

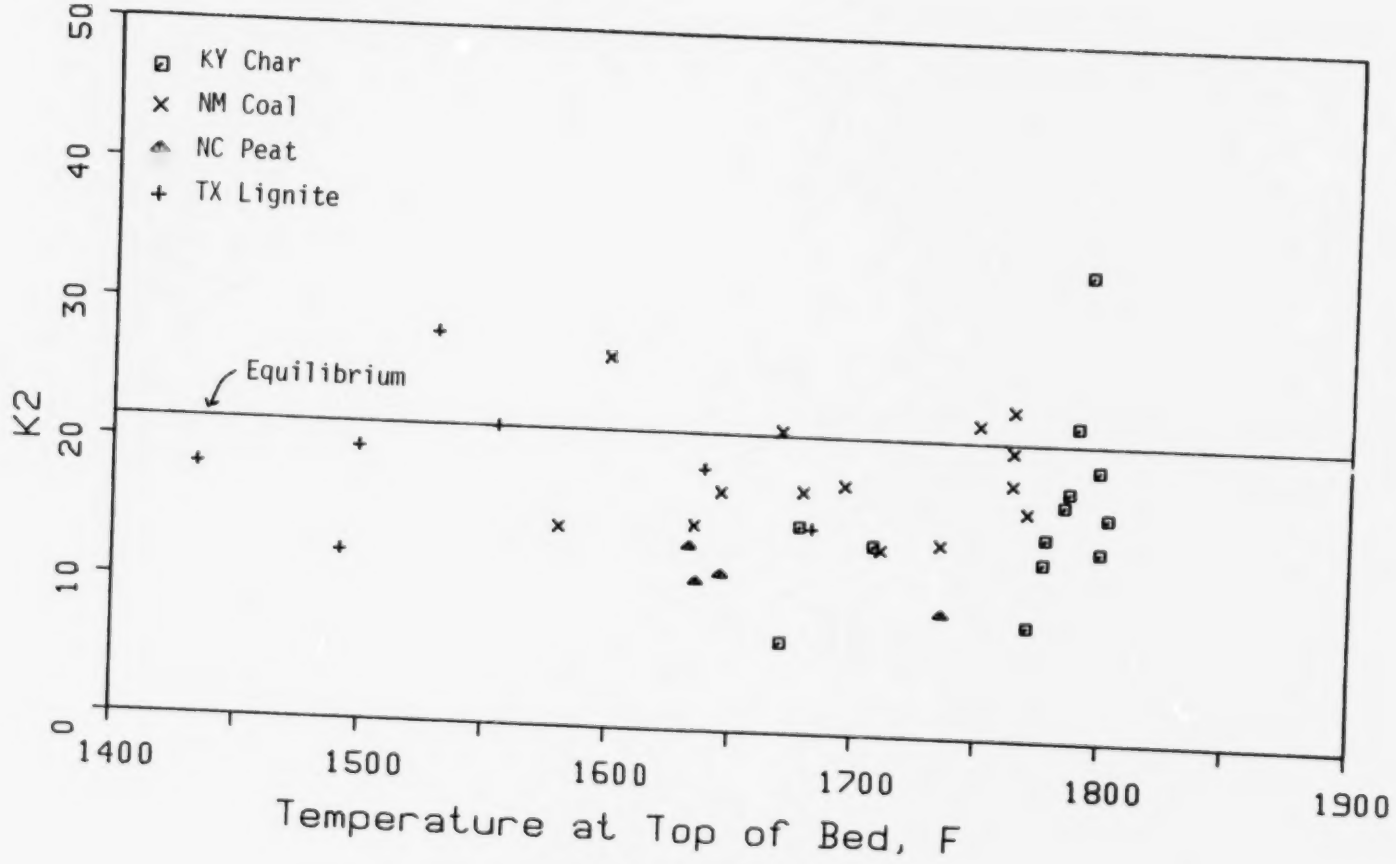

FIGURE 27

EXPERIMENTAL VALUES OF K_2 VERSUS TEMPERATURE

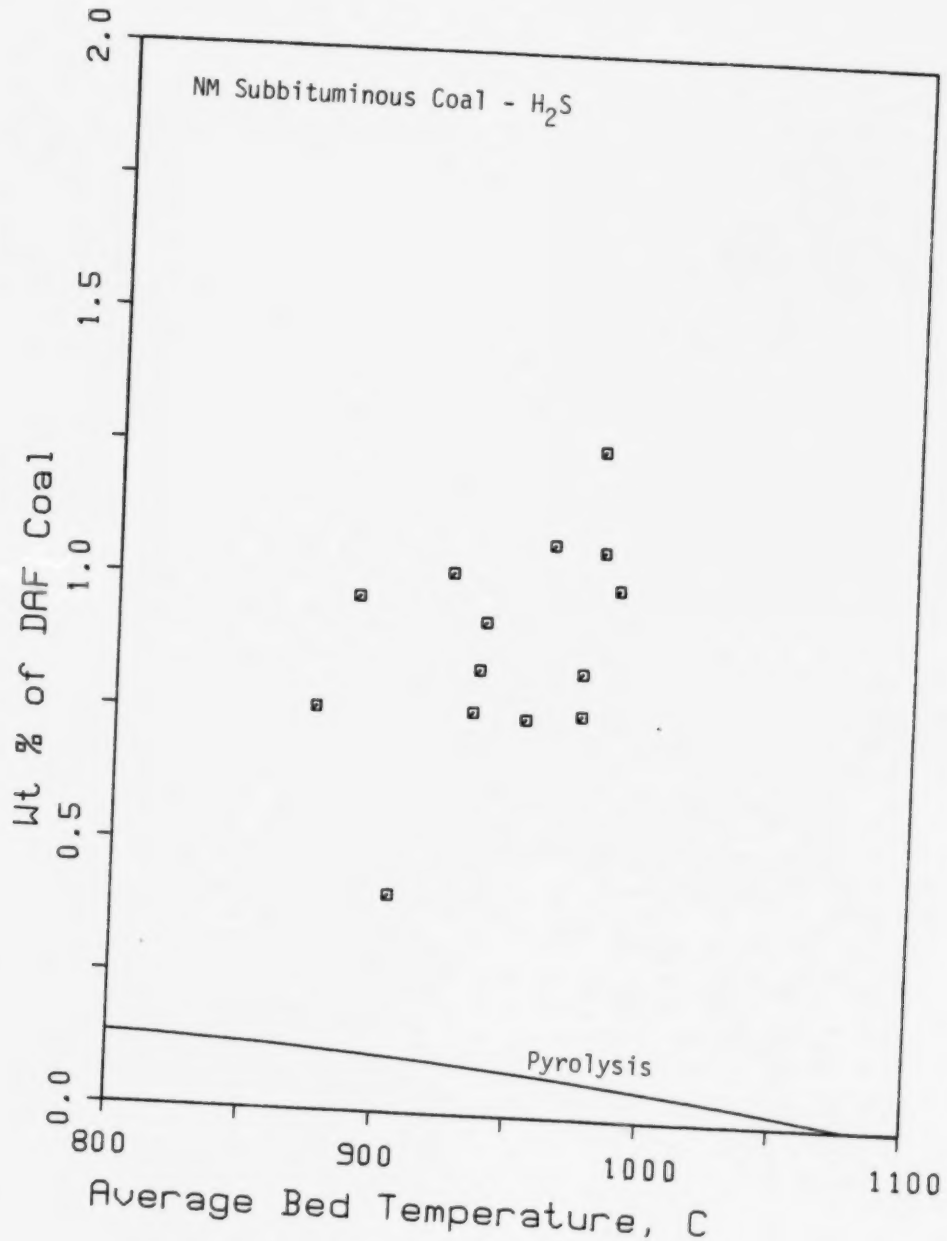

FIGURE 18

HYDROGEN SULFIDE PRODUCTION RATES FROM PYROLYSIS AND GASIFICATION OF NM COAL

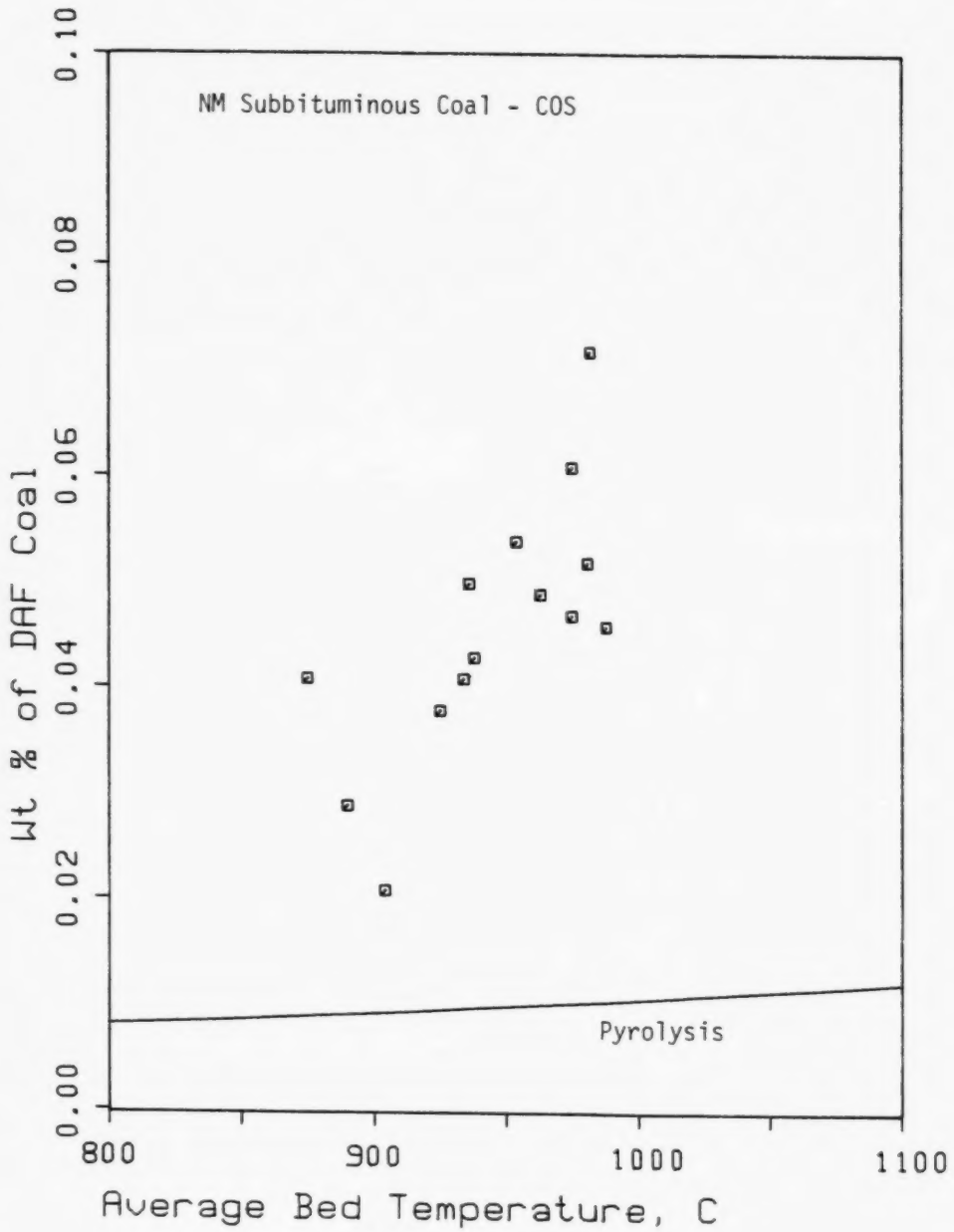

FIGURE 19

CARBONYL SULFIDE PRODUCTION RATES FROM PYROLYSIS AND GASIFICATION OF NM COAL

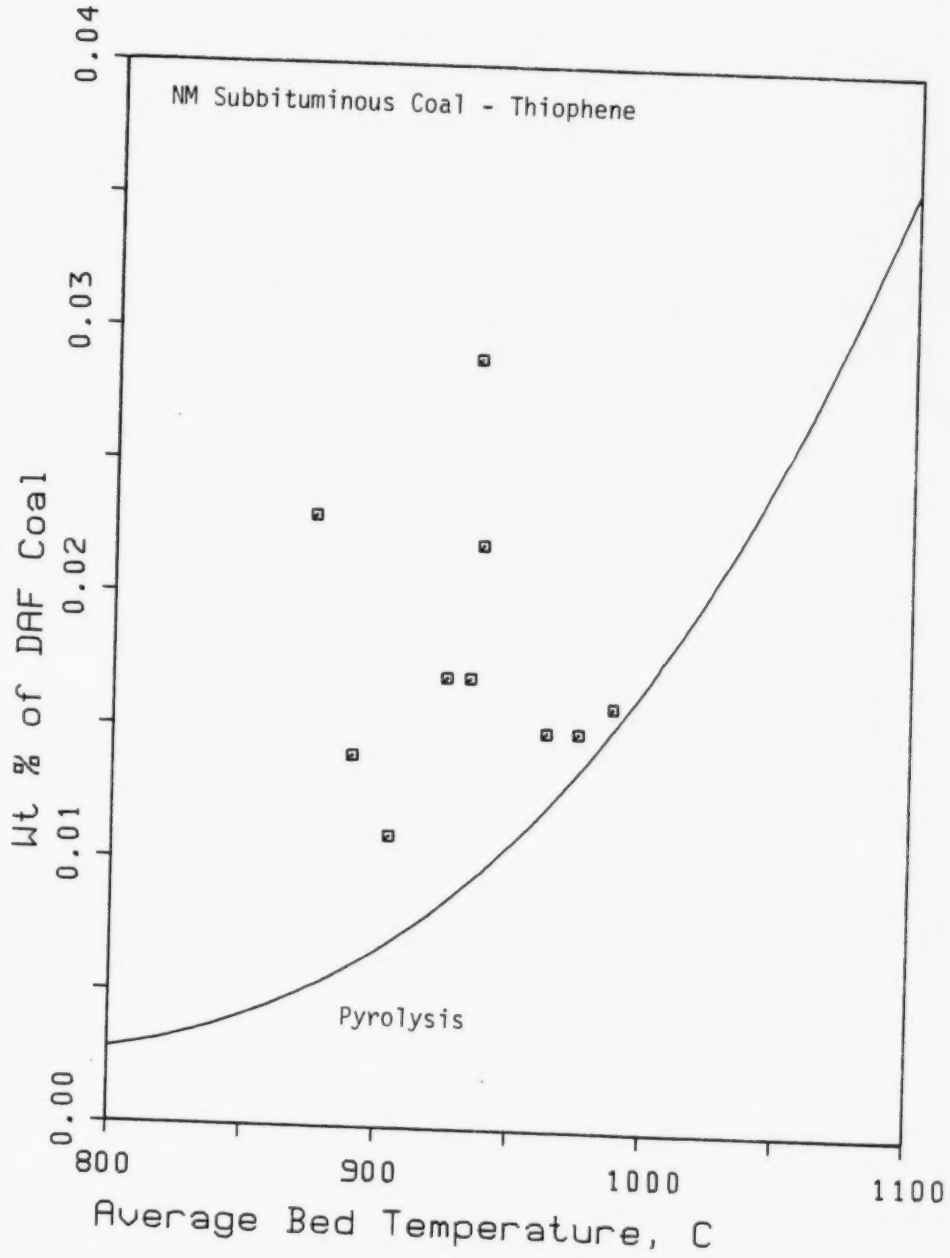

FIGURE 20

THIOPHENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

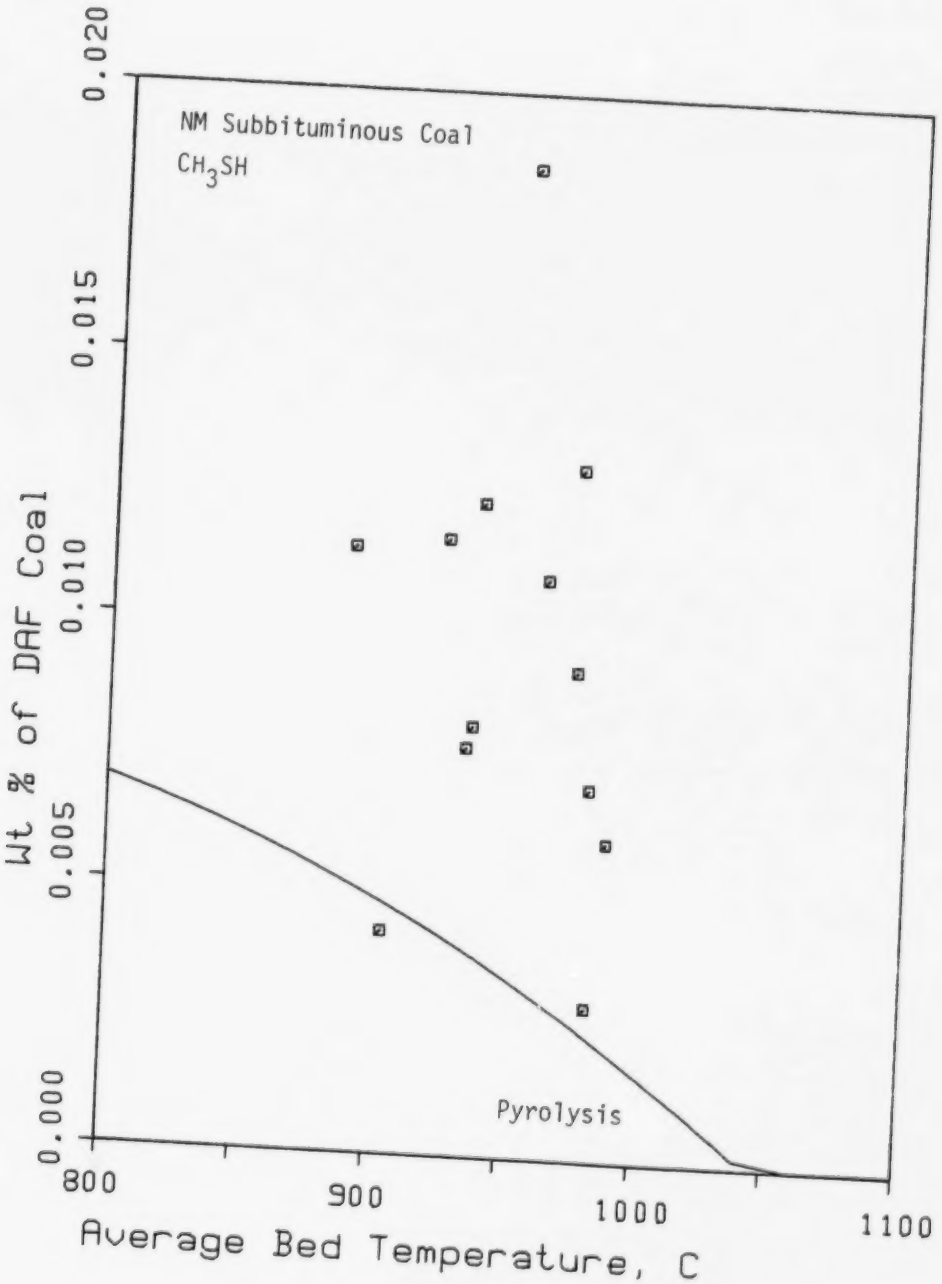

FIGURE 21
METHYL MERCAPTAN PRODUCTION RATES FROM PYROLYSIS AND GASIFICATION OF NM COAL

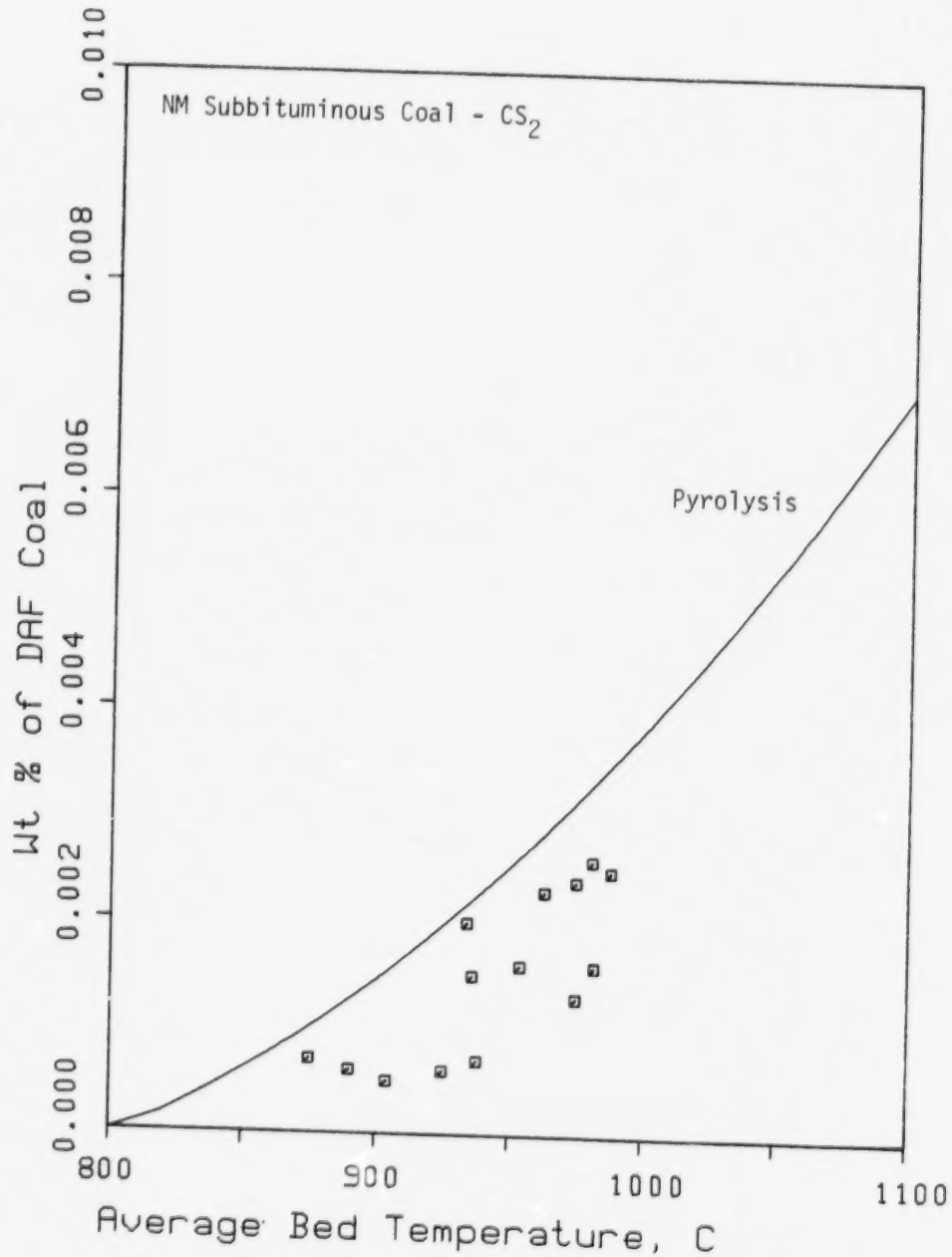

FIGURE 22

CARBON DISULFIDE PRODUCTION RATES FROM PYROLYSIS AND GASIFICATION OF NM COAL

TABLE 16
ALIPHATIC HYDROCARBON PRODUCTION RATES
Grams per kilogram of DAF feed

Run	Temp* (°F)	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4H_8	C_4H_{10}
NM Coal								
GO-66	1799	136	7.0	10.9	---	---	---	---
GO-68B	1787	89	9.4	15.7	---	---	---	---
GO-69B	1720	113	9.5	13.1	---	---	---	---
GO-72	1797	112	12.6	18.1	---	---	---	---
GO-74	1659	63	5.0	7.2	2.6	1.2	---	0.6
GO-76	1712	93	13.0	19.3	5.7	0.9	---	1.0
GO-78	1606	123	8.5	11.0	4.5	2.0	---	1.1
GO-79	1748	112	9.0	10.6	4.2	1.3	---	2.8
GO-80B	1765	98	14.7	13.5	6.0	1.8	2.0	1.1
GO-81B	1717	90	12.8	14.7	6.5	2.3	2.6	1.2
GO-81D	1633	96	12.9	13.2	7.6	2.3	3.0	1.4
GO-82	1810	103	13.9	14.2	5.2	1.6	2.4	1.2
GO-84	1697	92	13.4	15.7	6.2	1.9	2.5	1.1
GO-85B	1786	103	14.7	17.3	5.5	1.5	2.4	0.8
NC Peat								
GOP-1	1730	78	12.0	12.0	4.4	1.1	---	0.8
GOP-3	1582	80	12.5	1.7	---	---	---	---
GOP-4B	1607	65	8.6	13.8	15.7	4.1	---	2.9
GOP-6	1632	76	10.9	14.1	6.9	2.7	---	1.2
TX Lignite								
GOL-2	1644	75	10.6	15.2	5.4	2.0	3.2	0.9
GOL-3	1699	85	10.9	17.9	4.1	1.6	2.3	1.0
GOL-4	1546	78	7.8	11.8	4.9	2.1	2.9	0.9
GOL-5	1527	97	5.3	9.7	3.6	1.8	2.0	0.7
GOL-6	1536	77	5.9	7.8	3.3	1.3	1.9	0.5
GOL-7B	1520	71	7.0	11.3	2.7	1.1	1.6	0.7
GOL-11	1463	64	3.6	7.0	2.6	1.4	1.6	0.5

* volume-weighted average bed temperature

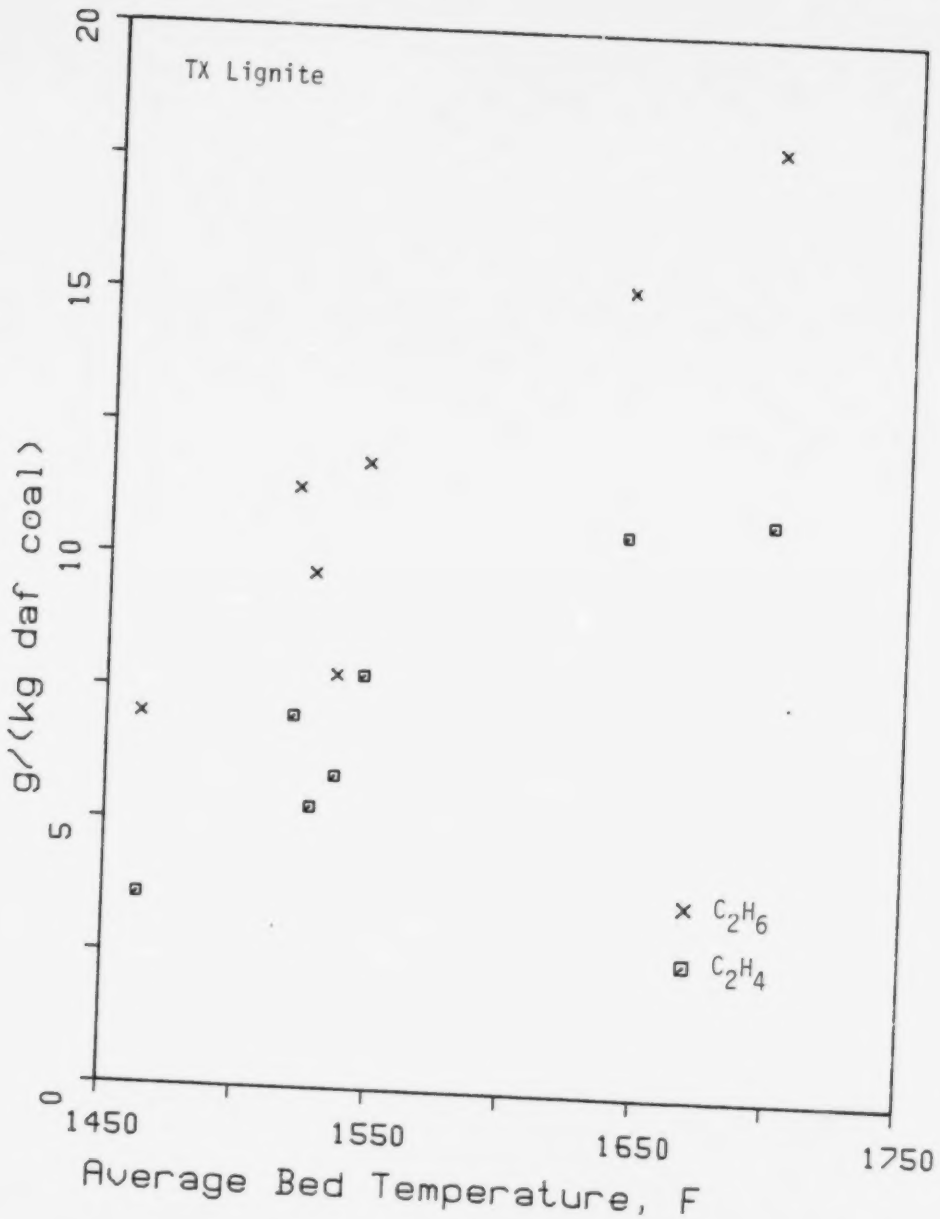

FIGURE 23

C_2H_4 AND C_2H_6 PRODUCTION RATES FOR TX LIGNITE

Figures 24-29 plot the production rates of aliphatic hydrocarbons from both gasification and pyrolysis (Gilman, 1983) of New Mexico coal. The data are plotted versus the average bed temperature for gasification, and against the peak temperature for pyrolysis. Although there is considerable scatter in the plots, it appears that the alkene production rates show reasonable agreement between gasification and pyrolysis. The alkanes, however, seem to show greater production levels for gasification than for pyrolysis. The differences may be explained by the heating rates involved in the two processes -- slow heating rates for the pyrolysis studies, and much faster heating rates during gasification. Future pyrolysis work, using a laminar flow reactor capable of fast heating rates, is expected to provide improved data for comparisons with gasification data.

Aromatic hydrocarbon production rates are listed in Table 17. No aromatics were detected from the gasification of the Kentucky char. For the other feedstocks, no major differences in aromatic hydrocarbon rates could be noted.

Comparisons of aromatic production rates from gasification and pyrolysis for New Mexico coal are shown in Figures 30-32. The pyrolysis data shown are those of Gilman. For all species shown, the production rates are greater from gasification. The reasons for the differences are unclear, but are probably related to the different heating rates as discussed previously.

Wastewater--

Water samples were collected from both the PCS tank and from the sample train. Samples from the PCS tank were collected after the completion of a run, and thus represent the accumulation of material during the times of startup, steady state operation, and shutdown. Water samples labeled as trap water were taken from the sample train, either from the condensate trap or from the cold tar trap after its removal at the end each run (see Figure 11). Trap water samples from either location of the sample train were essentially identical, and were from the steady state portion of each run. Both the PCS tank water and the sample train condensate water were analyzed as described in the analytical methods section of this report.

Table 18 shows typical results of analyses of the PCS water and trap water for the New Mexico coal runs, and is intended to show the general levels of concentrations as well as the differences between the two kinds of samples. As expected, the PCS values were lower than those for the trap water. The analyses of the PCS tank water are useful for general information; however, the sample train condensate analyses should be used for quantitative data. All further results to be presented are for the sample train trap water.

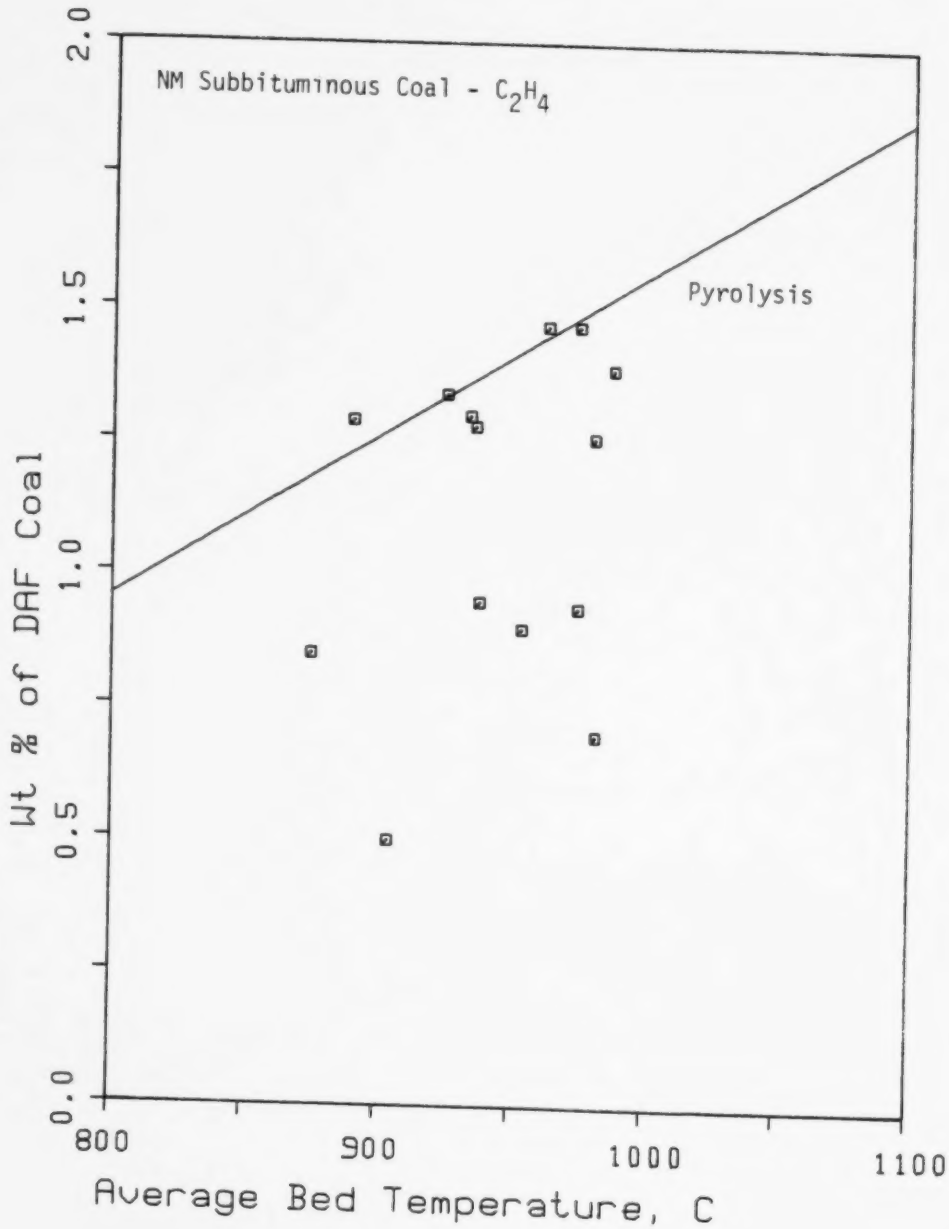

FIGURE 24

ETHYLENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

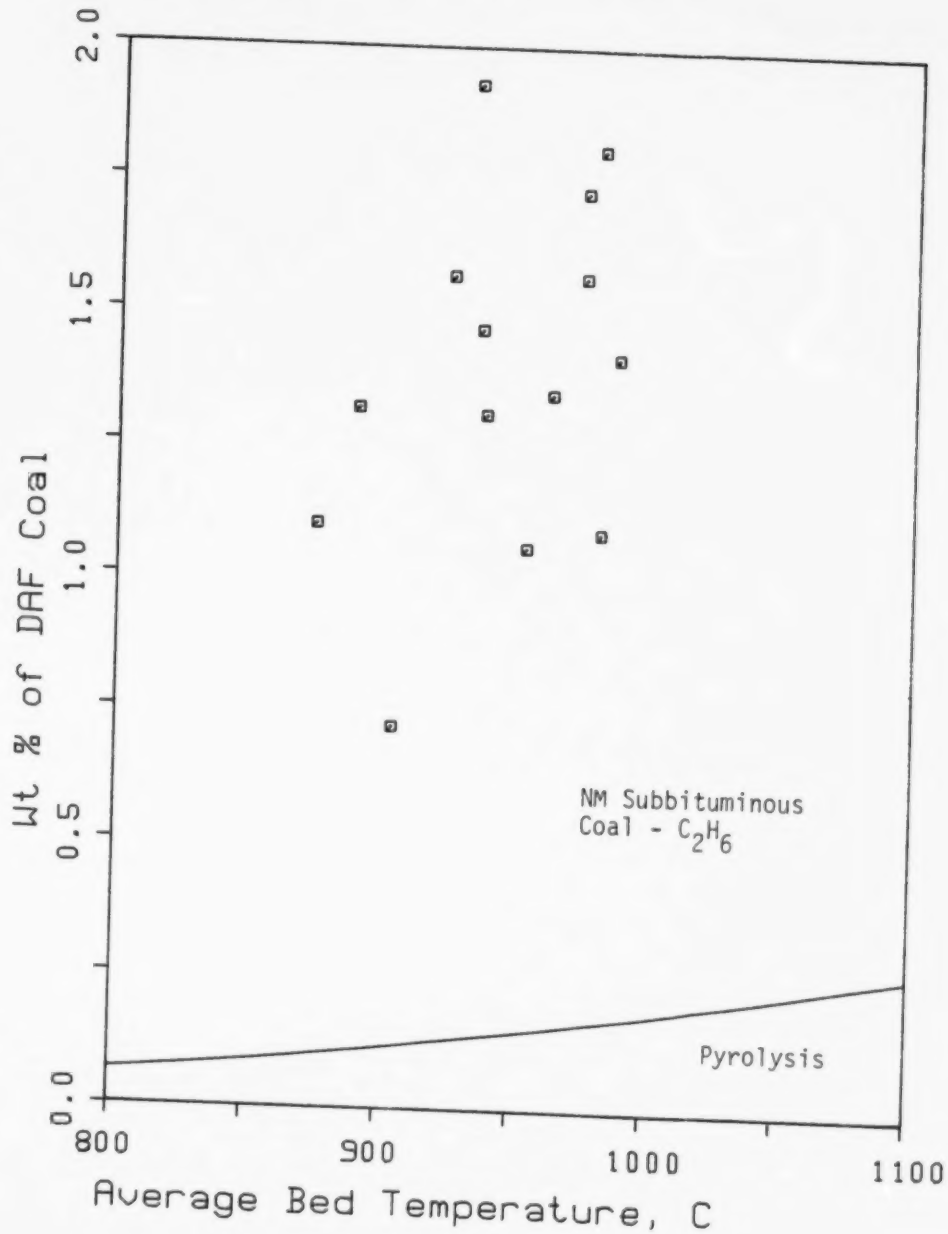

FIGURE 25

ETHANE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

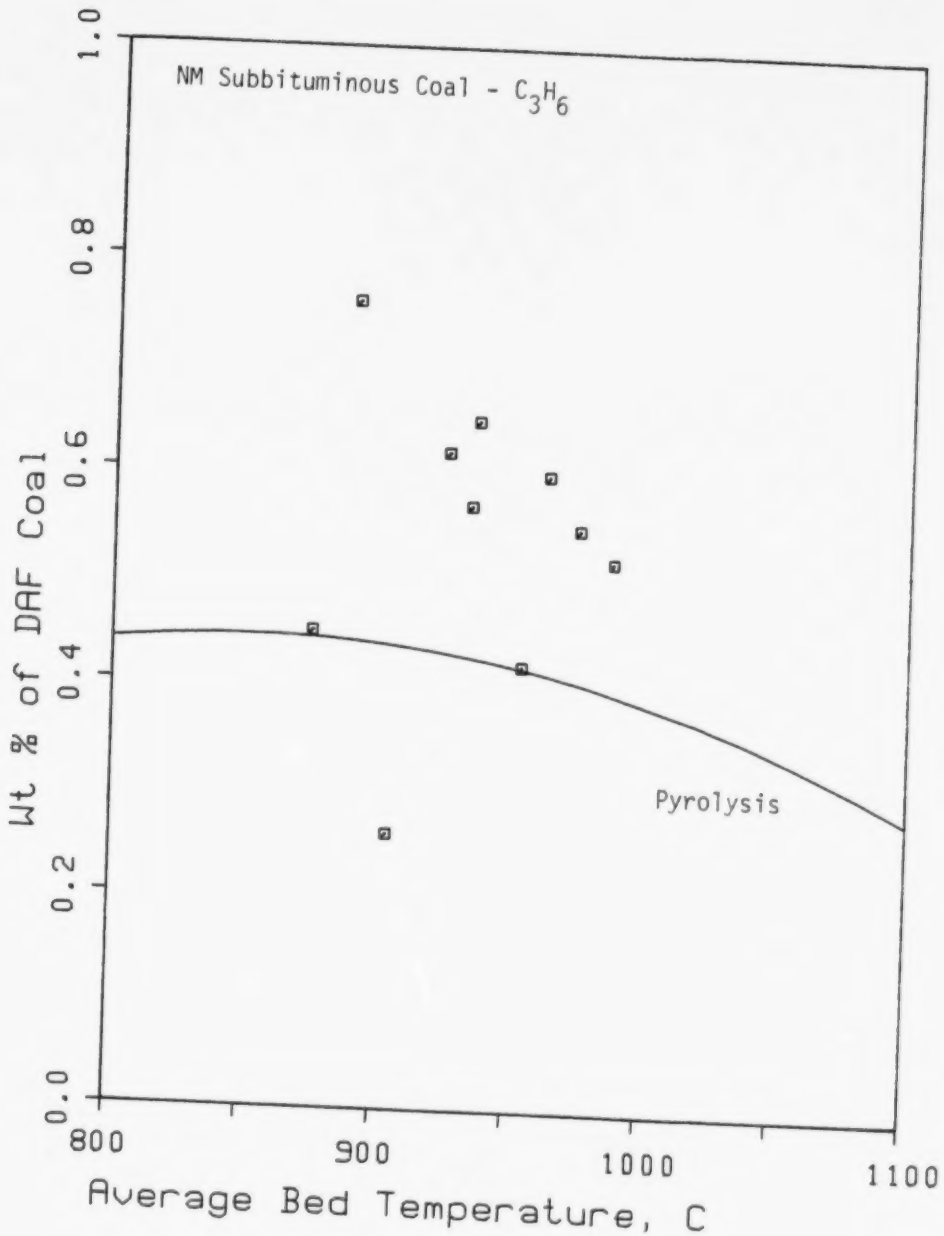

FIGURE 26

PROPENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

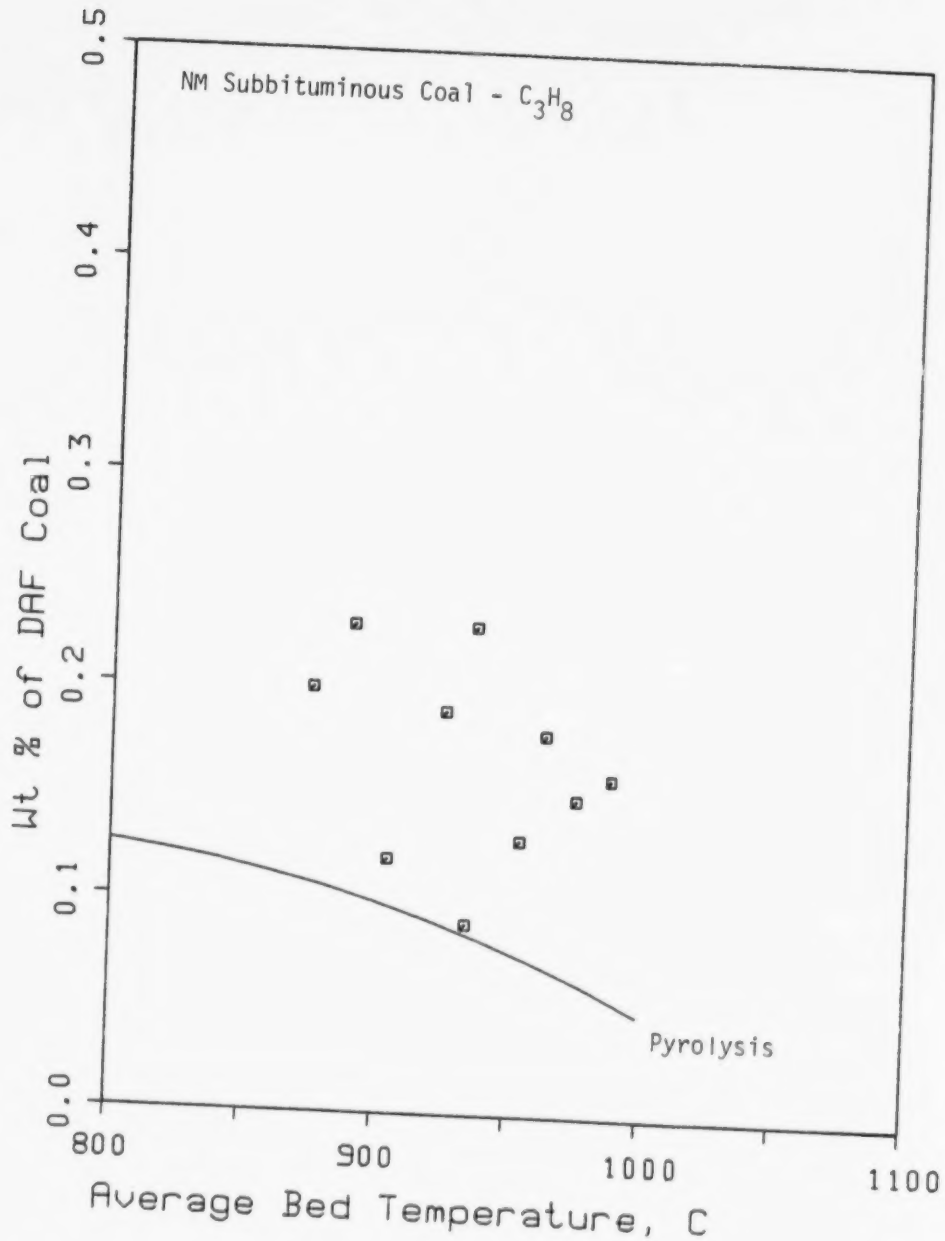

FIGURE 27

PROPANE PRODUCTION RATES FROM PYROLYSIS AND GASIFICATION OF NM COAL

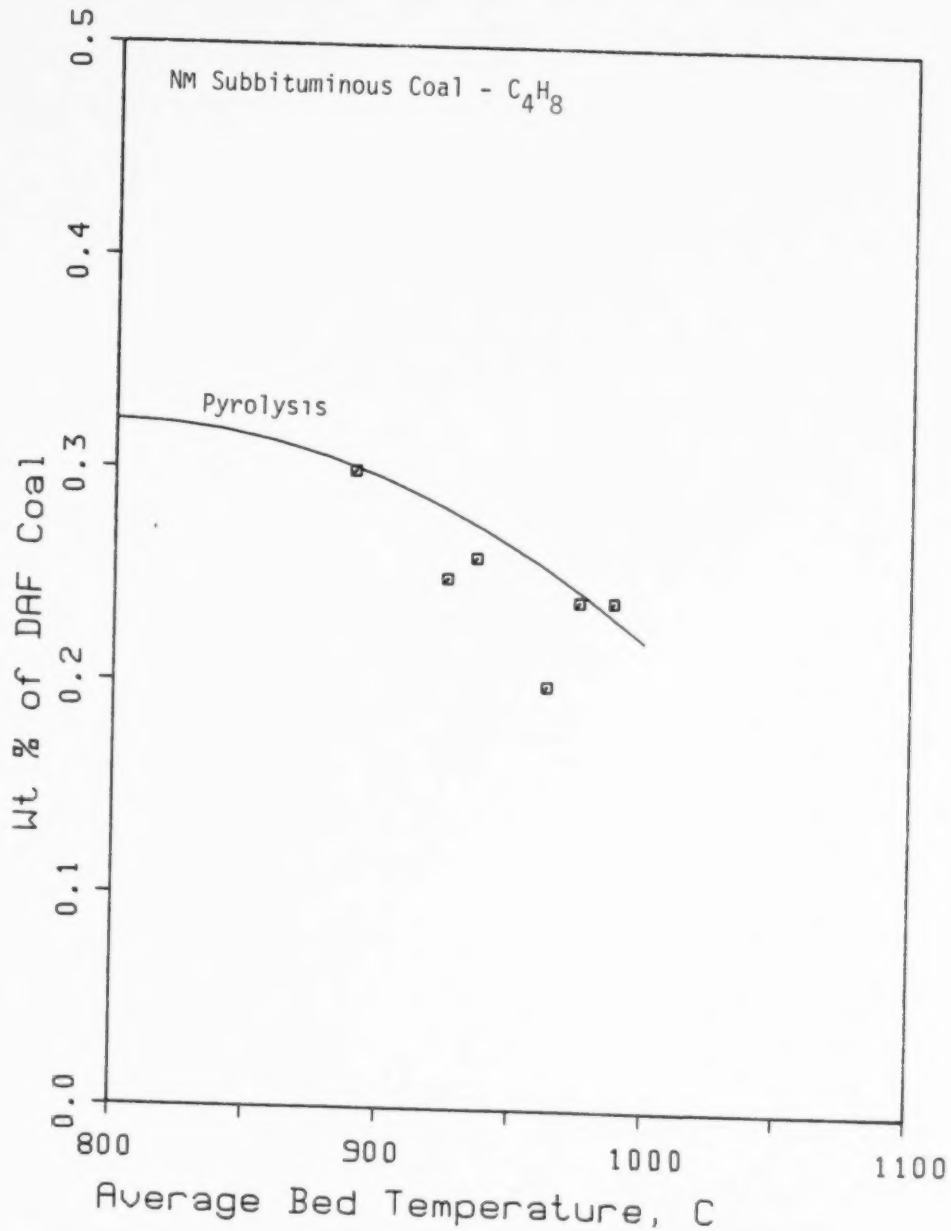

FIGURE 28

BUTENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

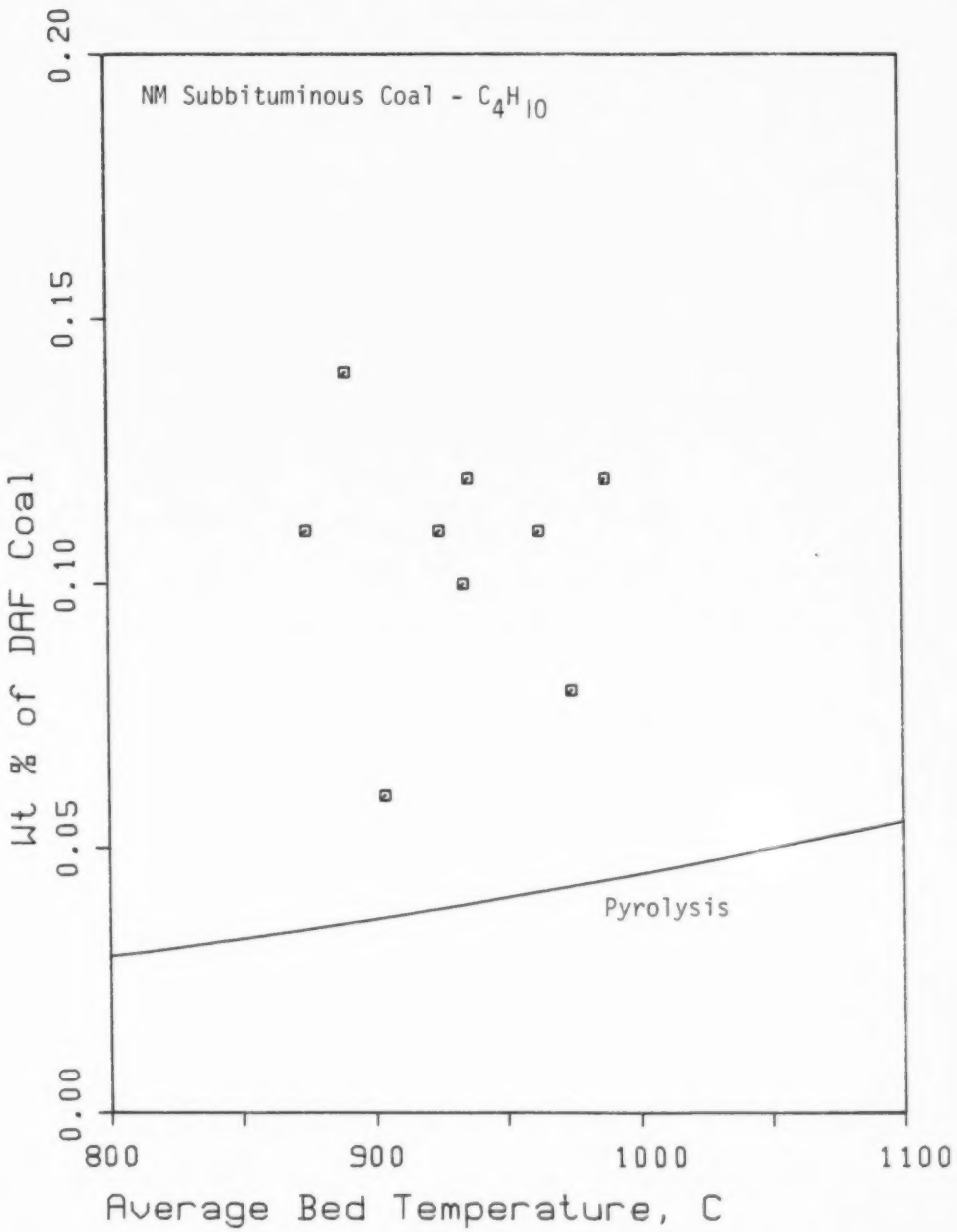

FIGURE 29

BUTANE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

TABLE 17
AROMATIC HYDROCARBON PRODUCTION RATES
Grams per kilogram of DAF feed

Run	Temp* (°F/°C)	Benzene	Toluene	Xylenes+	Ethylbenzene
NM Coal					
GO-66	1799/982	8.3	8.0	---	---
GO-68B	1787/975	---	---	---	---
GO-69B	1720/938	6.5	2.5	0.61	0.10
GO-72	1797/981	---	---	---	---
GO-74	1659/904	4.2	6.3	---	---
GO-76	1712/933	5.1	8.0	---	---
GO-78	1606/874	1.3	3.0	---	---
GO-79	1748/953	---	---	---	---
GO-80B	1765/963	6.0	1.8	---	---
GO-81B	1717/936	6.0	2.7	0.62	0.15
GO-81D	1633/889	6.8	4.1	0.67	0.12
GO-82	1810/988	12.1	2.5	0.30	0.07
GO-84	1697/925	4.8	1.7	0.32	0.08
GO-85B	1786/974	5.2	1.4	0.46	0.09
NC Peat					
GOP-1	1730/943	3.6	3.3	---	---
GOP-3	1582/861	4.7	8.0	---	---
GOP-4B	1607/875	6.2	2.5	---	---
GOP-6	1632/889	5.7	2.4	---	---
TX Lignite					
GOL-2	1644/896	5.6	2.2	0.86	0.20
GOL-3	1699/926	4.9	1.6	0.56	0.14
GOL-4	1546/841	3.0	1.3	0.53	0.17
GOL-5	1527/831	4.2	1.4	0.73	0.16
GOL-6	1536/836	3.7	1.3	0.56	0.13
GOL-7B	1520/827	4.0	1.1	0.39	0.09
GOL-11	1463/795	3.5	1.3	0.43	0.11

* volume-weighted average bed temperature
\+ sum of m-xylene, p-xylene, and o-xylene

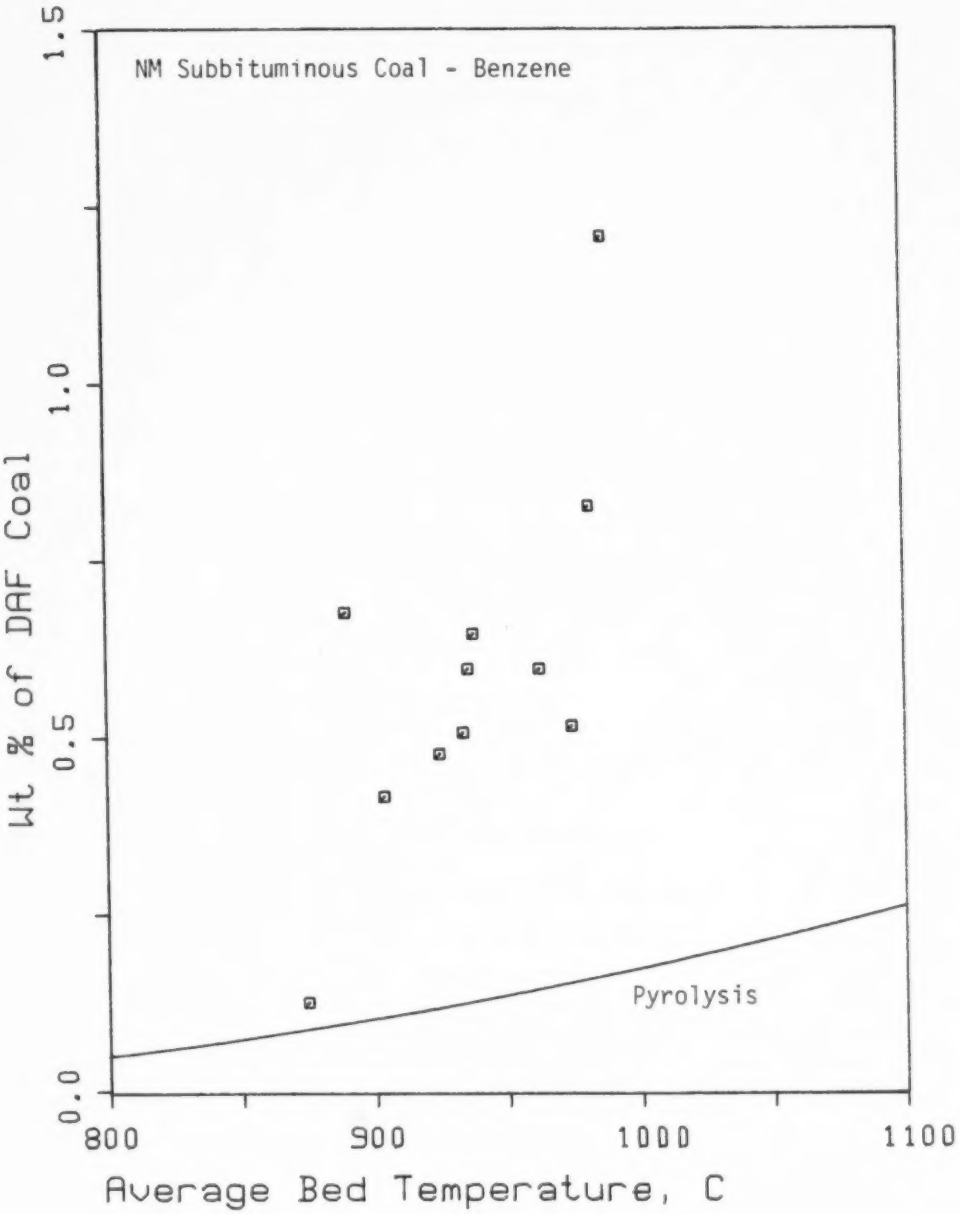

FIGURE 30

BENZENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

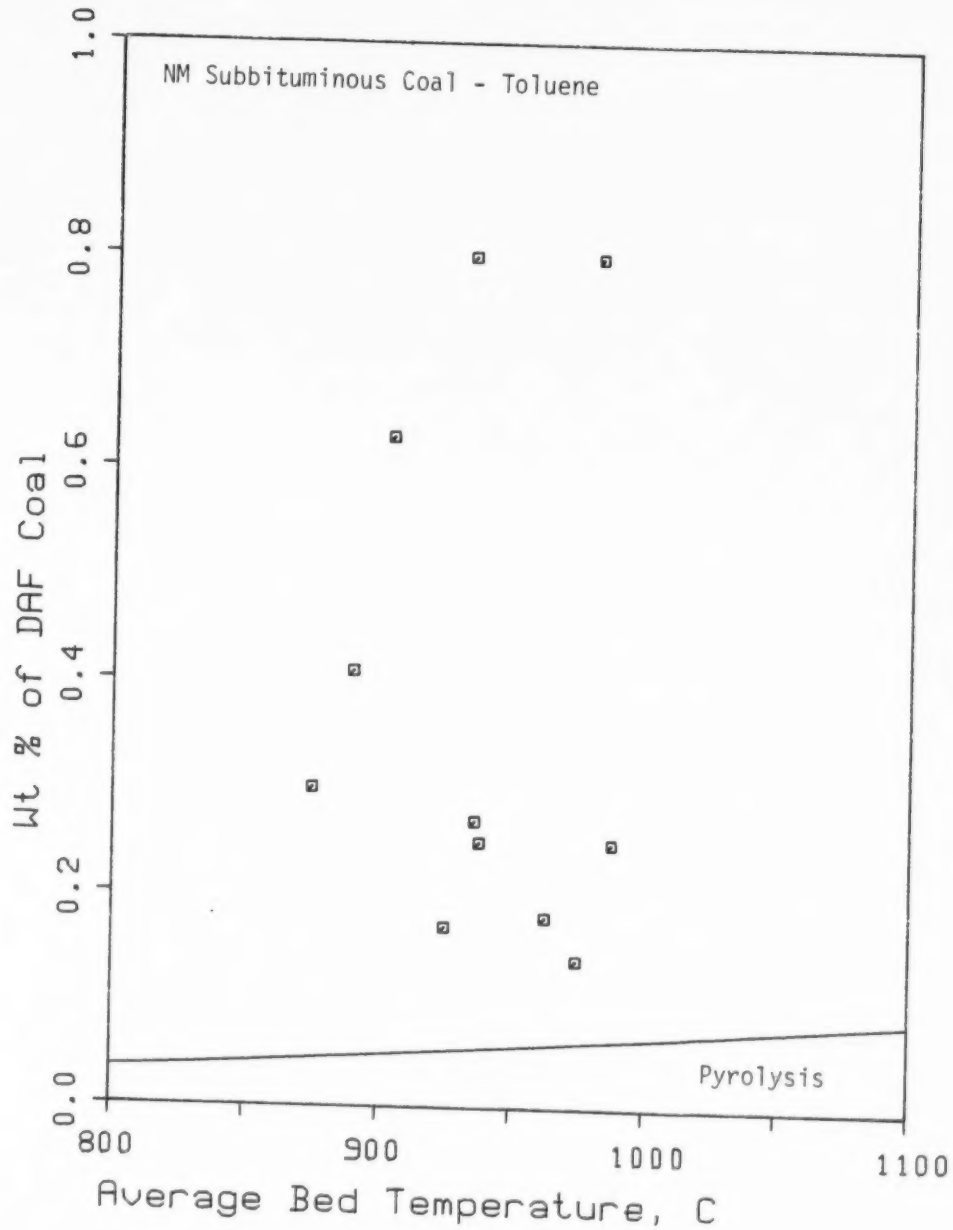

FIGURE 31

TOLUENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

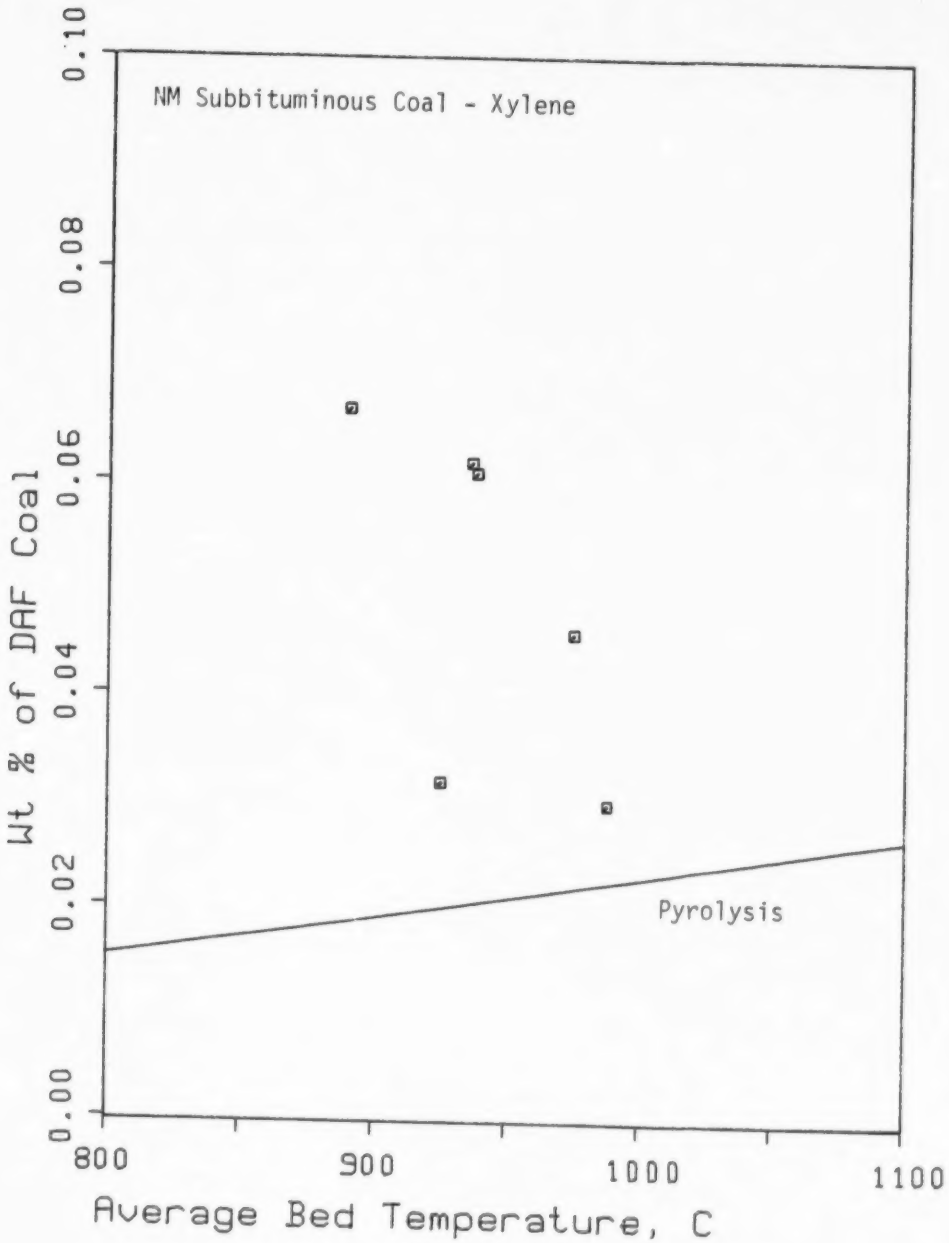

FIGURE 32

XYLENE PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

TABLE 18
COMPARISON OF PCS AND TRAP WATER ANALYSES FOR NM COAL
mg/l except pH

	PCS water	Trap water
Ammonia	700	6,000
Carbon	725	3,200
Chloride	20	40
COD	1,500 - 3,000	6,000 - 10,000
Cyanate	500	2,000 - 5,000
Cyanide	45	25 - 200
Fluoride	5	10
Nitrogen	600	6,000
pH	7.7	8.5
Phenolics	200 - 400	600 - 1,100
Sulfate	35	40 - 300
Sulfite	15	40
Thiocyanate	50	250
TOC	400 - 600	2,600
TVC	325	1,500

Table 19 summarizes the results of trap water analyses for each feedstock. An interesting result was that the char wastewater was slightly acidic, while the wastewaters from the other feeds were slightly basic. The reason for this difference was not determined, although ammonia produced in the devolatilization of the raw coals may be a factor. In addition, the water from char gasification runs was more dilute in nitrogen species. Comparison of the three raw coal feedstocks showed little difference except for dissolved carbon (Carbon, COD, TOC, and TVC). The higher dissolved carbon values for the peat are believed to be mostly in the form of phenols. This is consistent with several additional analyses which indicate that the gasification of peat produces relatively higher amounts of phenols and other acidic compounds than does the gasification of coal and lignite.

TABLE 19
SAMPLE TRAIN TRAP WATER SPECIES CONCENTRATIONS

	Range of Results, mg/l except pH			
	KY Char	NC Peat	NM Coal	TX Lignite
pH	6.3 - 7.0	8.2 - 8.9	8.0 - 8.7	8.1 - 8.4
Nitrogen	470 - 650	4700 - 9800	5400 - 8200	4500 - 8500
Ammonia	900 - 1300	3900 - 8000	3900 - 7400	5500 - 6100
Cyanide		37 - 75	10 - 270	50 - 180
Cyanate		600 - 1770	360 - 7200	3200 - 6000
Thiocyanate		230 - 430	100 - 700	254
Sulfate	6 - 40	2 - 20	10 - 300	3 - 7
Sulfite	6 - 40	6 - 60	10 - 70	9 - 18
Sulfide			830 - 840	30 - 60
Chloride	5 - 50	50 - 180	10 - 130	8 - 22
Fluoride	8 - 60	12 - 120	6 - 30	4 - 15
Carbon		10400 - 18000	2500 - 7500	3200 - 4400
COD		30500 - 39600	950 - 10300	10700 - 13300
TOC		9300 - 13600	1200 - 6800	2400 - 3700
TVC		2500 - 11000	900 - 2400	1000 - 2400

Table 20 shows the data of Table 19 normalized to show the rate of production of the various species in milligrams per kilogram of coal fed to the gasifier. The results for coal, peat, and lignite are nearly the same, except for those pertaining to dissolved carbon (carbon, COD, TOC, and TVC), for which peat had the highest values. As previously noted, the nitrogen species levels were lower for the char wastewater than for the others. The various wastewater species production rates were examined to determine the effect of the reactor temperature. No clear trends were observed, indicating that for the temperature range covered the reactor temperature has little effect on the production rates of wastewater species.

TABLE 20
SAMPLE TRAIN TRAP WATER SPECIES PRODUCTION RATES

	Range of Results, g/kg feed			
	KY Char	NC Peat	NM Coal	TX Lignite
Nitrogen	0.4 - 0.9	3.5 - 6.5	5 - 9	3.8 - 5.5
Ammonia	1.2 - 1.6	2.8 - 5.2	4.0 - 8.3	4.1 - 4.6
Cyanide		0.02 - 0.04	0.01 - 0.3	0.03 - 0.13
Cyanate		0.3 - 1.5	0.4 - 6.8	2.7 - 4.5
Thiocyanate		0.1 - 0.5	0.1 - 0.7	0.2
Sulfate	0.004 - 0.05	0.001 - 0.023	0.02 - 0.28	0.002 - 0.005
Sulfite	0.003 - 0.05	0.004 - 0.07	0.007 - 0.08	0.008 - 0.013
Sulfide			0.9 - 1.0	0.02 - 0.05
Chloride	0.006 - 0.03	0.03 - 0.14	0.01 - 0.2	0.007 - 0.016
Fluoride	0.01 - 0.06	0.01 - 0.06	0.004 - 0.02	0.004 - 0.011
Carbon		7 - 12	2 - 9	2 - 4
COD		21 - 36	1 - 11	8 - 10
TOC		6 - 11	1 - 7	2 - 3
TVC		2 - 7	1 - 3	1 - 2

The results given in Tables 19 and 20 agree reasonably well with data presented in studies of wastewater from fixed and fluidized bed gasification (Forney et al., 1975; Nakles and Massey, 1977; Singer et al., 1981; Neufeld et al., 1981; Cleland and Gangwal, 1981; Hung et al., 1982). The differing processes studied and coals utilized can account for the variances in the measurement of wastewater parameters.

In addition to the routine analysis results discussed here, detailed analyses were performed for phenolic and polynuclear aromatic compounds in the wastewater. The methods and results of these additional analyses are not included in this report, as they are described in the works of Hubik and Ferrell (1982) and Bradley (1983).

Tar

Gasification of the raw coals yielded substantial amounts of tar. The tar production rate from New Mexico coal was great enough to cause modification of the PCS system, as previously discussed. Following the modifications, the facility has been operated without shutdowns forced by PCS system clogging.

The tar evolution rate is determined from cold trap measurements. The trap is operated for one hour during steady-state to obtain samples of tar, water, and particulates. The total tar rate is determined from the cold trap tar/water ratio and the total water effluent rate as

$$\text{Tar Rate} = (H_2O \text{ Mass Flow Rate}) \frac{\text{mass of tar}}{\text{mass of } H_2O} \quad (7)$$

The tar rates calculated in this manner are tabulated in Table 21. Only those runs for which the cold trap was successfully operated are included. Tar rates, given as fractions of the dry ash-free feed coal, are also listed. The tar rates are plotted in Figure 33 versus the average bed temperature. The data show no correlation with temperature. The reason for the apparent lack of temperature dependence is probably due to tar formation being a freeboard process. In the freeboard region of the reactor, the temperature varied little from run to run, despite large variations in the fluidized bed temperature. The fluidized bed temperature thus has little effect on the tar production rate. Comparison of the different feeds shows that the New Mexico coal has the greatest evolution rates of tar, with peat and lignite being roughly the same. The devolatilized bituminous coal yielded no measurable tar.

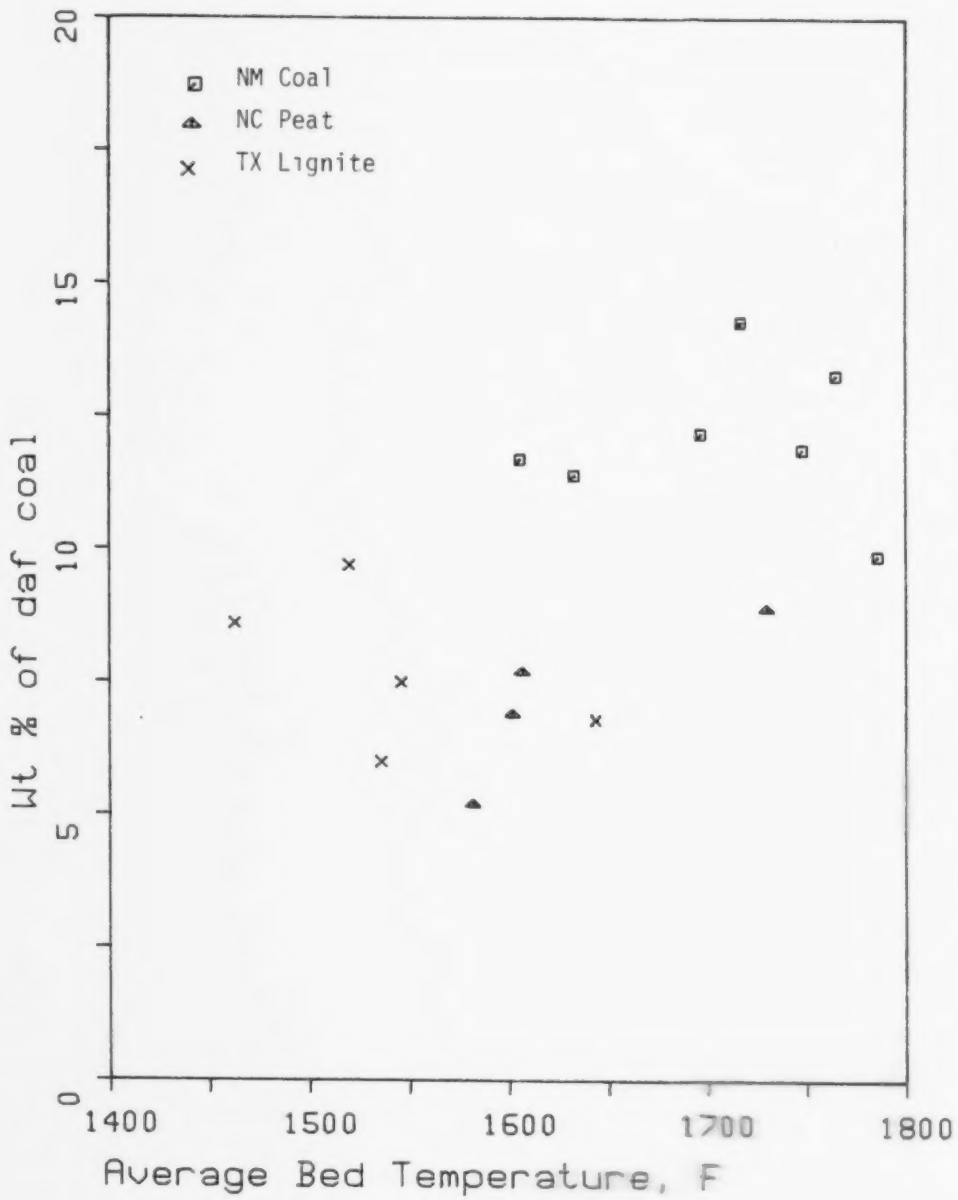

FIGURE 33
TAR YIELDS

TABLE 21
TAR YIELDS

	Temp,(°F)*	Tar Rate, lb/hr	Tar/(DAF Feed)	% C Loss
NM Coal				
GO-78	1606	3.40	0.117	13.0
GO-79	1748	3.47	0.119	13.4
GO-80B	1765	3.73	0.133	15.3
GO-81B	1717	4.92	0.143	15.9
GO-81D	1633	3.72	0.114	12.7
GO-84	1697	3.67	0.122	13.5
GO-85B	1786	3.30	0.099	10.7
NC Peat				
GOP-1	1730	2.30	0.089	11.3
GOP-2	1602	3.25	0.069	8.9
GOP-3	1582	2.14	0.052	6.7
GOP-4B	1607	3.25	0.077	9.8
TX Lignite				
GOL-2	1644	2.16	0.068	7.4
GOL-4	1546	2.86	0.075	8.2
GOL-6	1536	1.97	0.060	6.8
GOL-7B	1520	3.14	0.097	10.3
GOL-11	1463	3.24	0.086	9.6

* volume-weighted average bed temperature

The relative tar production rates for the three raw coal feedstocks measured in this study are consistent with the work of Solomon (1978,1981). Solomon suggested that tar consists of aromatic "monomers" released from the coal "polymer", and that the aromatic content of coals can be correlated with the fixed carbon content determined by proximate analysis. The proposed coal structures of Wender (1976) are consistent with this viewpoint (see Figure 1). Thus, greater tar yields would be expected from coals with greater aromaticity, which should be those with greater fixed carbon values. The New Mexico coal had the greatest tar production rate and also the greatest fixed carbon content (36%). The peat and lignite had similar tar rates, and also similar fixed carbon values -- 26% for the peat, and 23% for the lignite.

Carbon losses in the tar, calculated as percent of the feed carbon collected in the effluent tar, are listed in Table 21. The losses ranged from 7% to 16%, with the average being around 10%. Similar values were observed for all feeds, with the losses being slightly greater for the NM coal. The losses represent a significant portion of the feed carbon, and suggest the need for either further processing of the tar to recover products or tar recycle to the gasifier. Assumed tar

elemental compositions were used in the calculation of carbon losses. The compositions were estimated from the data of Gilman (1983) and Suuberg et al. (1979). Table 22 gives the assumed compositions used in the above calculations and in gasifier mass balance calculations.

TABLE 22
ESTIMATED TAR COMPOSITIONS, WT %

	NM Coal	NC Peat	TX Lignite
Carbon	85.0	80.0	80.0
Hydrogen	5.0	10.0	10.0
Oxygen	8.0	8.5	8.5
Nitrogen	0.5	0.5	0.5
Sulfur	1.0	0.5	0.5

Figure 34 plots the tar production rate for New Mexico coal from gasification and pyrolysis (Gilman, 1983). The gasification data points do not differ greatly from the curve for pyrolysis. If the gasification data points were plotted against freeboard temperature instead of average bed temperature, as perhaps they should be, the agreement would be even better This suggests that the greatest portion of the tar evolved during gasification originates in devolatilization.

Cyclone Solids

The cyclone separator is located just downstream of the gas exit from the reactor vessel. The solids collected accumulate in a receiving vessel. Following the completion of the run, the receiver is opened, and the solids are weighed and sampled. The flow rate of solids removed by the cyclone is estimated in the following manner:

$$\text{Cyclone Rate} = (\text{Feed Rate}) \frac{\text{Total cyclone solids}}{\text{Total coal fed}} \qquad (8)$$

Solids that pass through the cyclone are trapped in either the PCS tank or the PCS filters. The exit particulate rate is estimated from cold trap data. The particulate/water ratio is determined from cold trap measurements, and from the known total water effluent rate the particulate rate is calculated. The cyclone inlet particulate rate is the sum of the cyclone collection rate and the cyclone exit particulate rate. Overall cyclone efficiencies, defined as

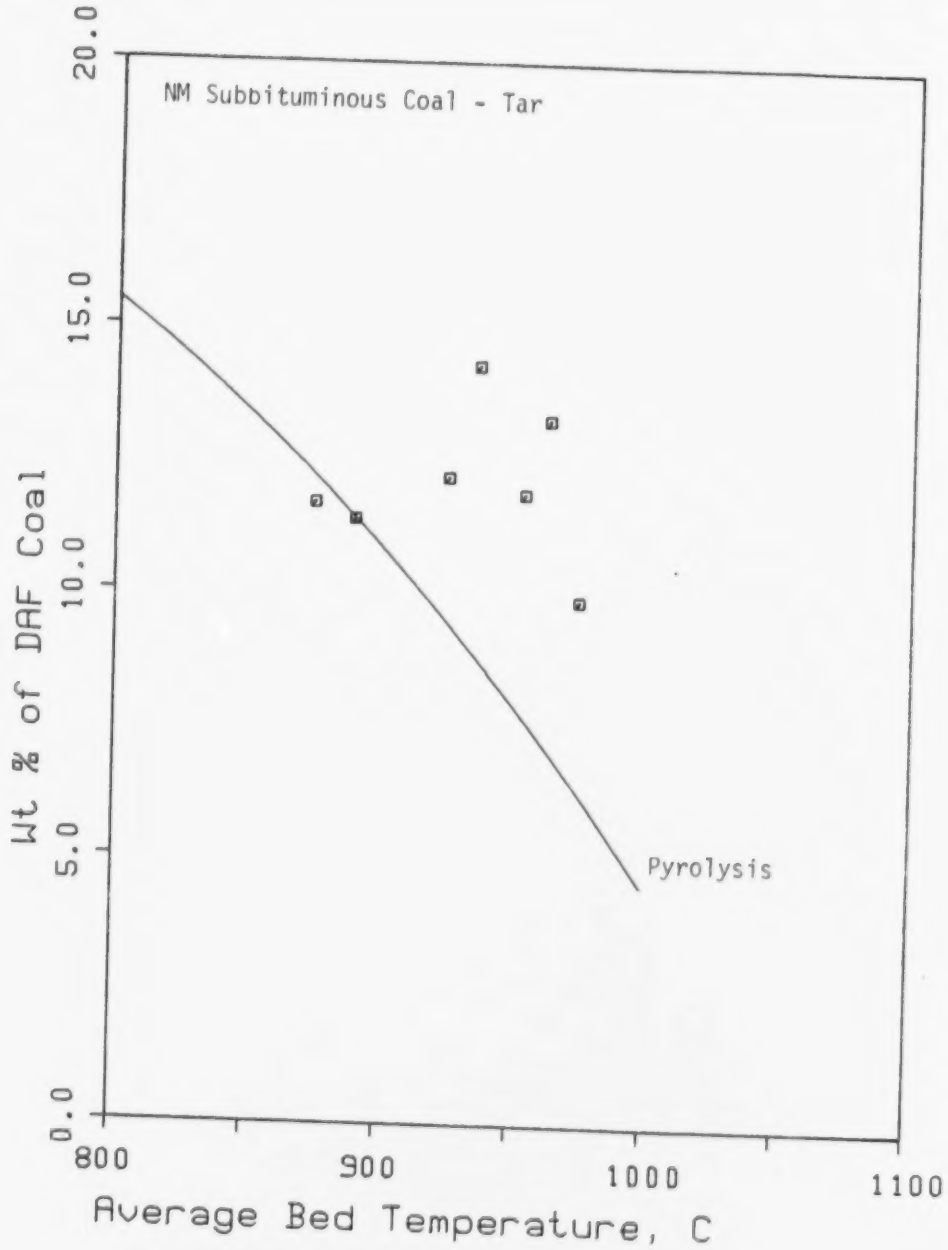

FIGURE 34

TAR PRODUCTION RATES FROM PYROLYSIS
AND GASIFICATION OF NM COAL

$$\text{Cyclone Efficiency} = \frac{\text{Cyclone collection rate}}{\text{Cyclone inlet rate}} \times 100 \qquad (9)$$

were calculated for all runs for which cold trap particulate data were available. The efficiency typically was 85-95% for runs in which the cyclone functioned properly.

For a large number of the New Mexico coal runs the cyclone did not function properly, as no solids were collected in the receiver vessel. The main cause was probably tar condensation preventing the solids from reaching the receiver. Tar condensation could easily occur during startup when the cyclone is below the condensation temperature, roughly 320°C. During steady-state operation, the cyclone inlet and outlet temperatures were typically 550°C and 500°C respectively, so that tar condensation was not a problem once steady-state was achieved.

Table 23 summarizes the inlet and outlet cyclone particulate rates and cyclone efficiencies. Also included in the table are the cyclone rates expressed as percents of the as-received coal feed, and the percent of the feed carbon that leaves the reactor in the cyclone solids. A significant fraction of the feed coal mass appears in the cyclone, averaging 10% for all the feedstocks. The carbon lost in the cyclone stream varied for the different feeds, with losses the greatest for peat and roughly the same for coal and lignite. The carbon loss was as high as 17% for the peat, indicating the desirability of fines recycle for this material.

TABLE 23
CYCLONE PERFORMANCE

	Inlet Rate, lb/hr	Collection Rate, lb/hr	% of Feed*	% C Loss*	Efficiency %
NC Peat					
GOP-1	3.28	3.06	8.9	14.2	93.3
GOP-2	5.23	4.93	7.6	13.1	94.3
GOP-3	4.02	3.52	7.0	10.5	87.6
GOP-4B	6.51	5.95	11.2	17.0	91.4
GOP-6	7.02	6.69	9.8	15.7	95.3
NM Coal					
GO-68B	--	4.45	7.0	3.7	--
GO-72	--	4.80	11.3	4.8	--
GO-74	--	3.15	4.7	2.7	--
GO-78	4.47	4.20	10.9	9.0	94.0
GO-79	4.61	2.22	11.0	7.1	48.2
GO-84	5.04	4.86	10.5	9.0	96.4
TX Lignite					
GOL-2	8.81	8.12	13.8	6.6	92.2
GOL-3	8.49	7.92	12.8	5.1	93.3
GOL-4	7.19	6.02	9.9	8.2	83.7
GOL-5	--	5.21	8.7	6.5	--
GOL-6	5.82	4.97	10.3	7.2	85.4
GOL-7B	8.79	6.53	16.4	9.8	74.3
GOL-11	8.03	6.86	12.0	8.0	85.4

* calculated from cyclone inlet rate when known, from cyclone collection rate otherwise.

Proximate and ultimate analyses were performed for all runs, with the range of results listed in Table 24 and complete results tabulated in Appendix 1.

TABLE 24
CYCLONE SOLIDS COMPOSITION

	Range of results, weight percents			
	KY Char	NM Coal	NC Peat	TX Lignite
Carbon	58 - 81	22 - 56	67 - 74	15 - 32
Hydrogen	0.2 - 0.8	0.7 - 4.0	1.6 - 2.5	0.6 - 2.2
Oxygen	1 - 20	2 - 13	9 - 12	1 - 10
Nitrogen	0.1 - 1.0	0.3 - 0.9	0.6 - 1.3	0.3 - 0.5
Sulfur	1.2 - 2.3	0.2 - 0.7	0.01 - 0.11	0.2 - 0.5
Ash	12 - 34	30 - 70	13 - 18	58 - 80

Spent Char

Samples of the spent char were taken from two locations following the completion of each run. For the majority of the runs, samples were taken from the top of the material in the char receiver. This sampling location was effective when the run produced a large amount of steady-state char. For the raw coal feedstocks, the amount of steady-state char was often much less than the amount of char produced during reactor startup, and representative char sampling was difficult. To avoid this problem, char samples were taken from the material left in the bed at shutdown for several of the peat, coal, and lignite runs. For a well-mixed fluidized bed, the bed and receiver samples should have matching compositions. The well-mixed assumption was checked by running proximate and ultimate analyses for several runs for both sampling points and comparing the results. Figures 35-37 compare the results of ash, carbon, and sulfur analyses ran on char samples from the bed and receiver. The two sample locations compare well with the exception of carbon and sulfur for one lignite run. For this run it is likely that the receiver sample was contaminated with the Kentucky char used for startup. In general the two locations compare well, indicating that the well-mixed solids assumption is probably valid.

The ranges of composition measured for the spent char for each feedstock are given in Table 25, and complete ultimate and proximate results are given in Appendix 1. In general, the char showed a lower carbon content and a greater ash content in comparison to the feed materials. An interesting exception was the NC peat, for which the char carbon content was often greater than the feed carbon content. The peat had a high content of oxygen which was depleted to a much greater extent than was carbon. Enrichment of the carbon content occurred as a result.

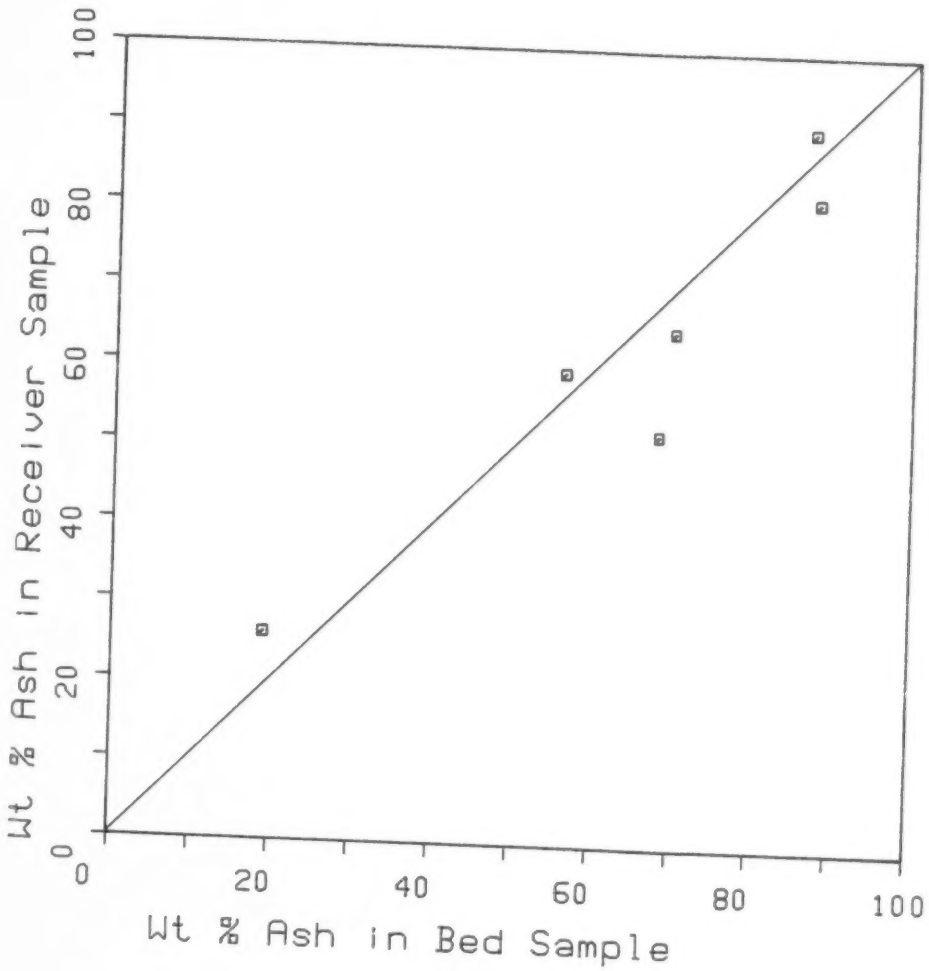

FIGURE 35

COMPARISON OF RECEIVER AND BED SOLIDS -- ASH CONTENT

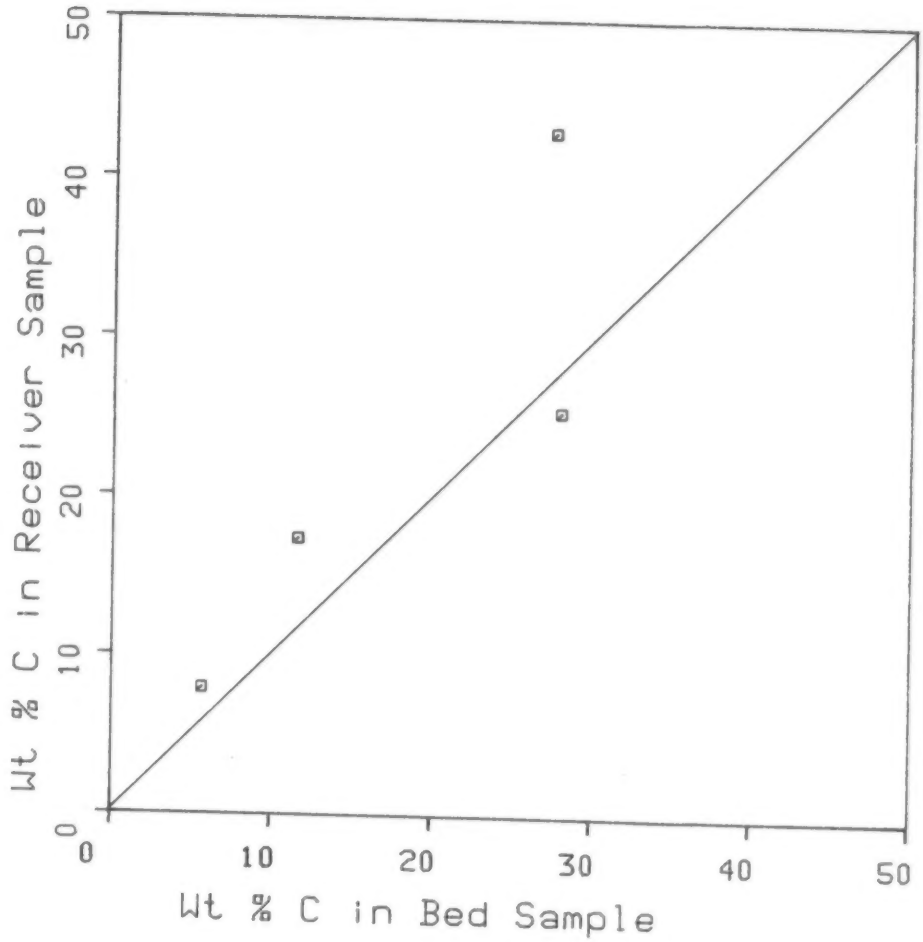

FIGURE 36

COMPARISON OF RECEIVER AND BED SOLIDS -- CARBON CONTENT

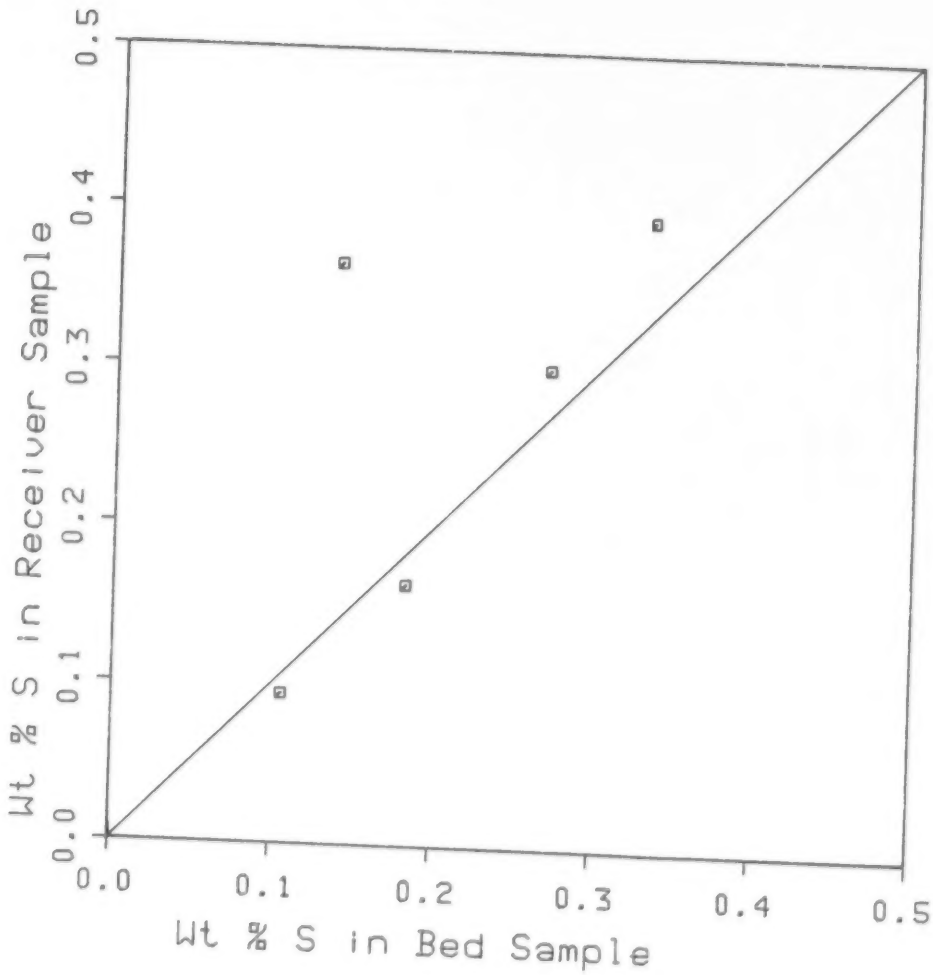

FIGURE 37
COMPARISON OF RECEIVER AND BED SOLIDS -- SULFUR CONTENT

TABLE 25
SPENT CHAR COMPOSITION

	Range of results, weight percents			
	KY Char	NM Coal	NC Peat	TX Lignite
Carbon	64 - 84	8 - 55	39 - 66	8 - 30
Hydrogen	0.2 - 0.5	0.3 - 1.8	0.6 - 1.5	0.5 - 1.8
Oxygen	0.2 - 7.1	0.3 - 14.0	1 - 20	1 - 14
Nitrogen	0.1 - 1.0	0.1 - 0.9	0.5 - 0.9	0.1 - 0.4
Sulfur	1.5 - 2.1	0.3 - 0.8	0.08 - 0.35	0.1 - 0.5
Ash	11 - 31	39 - 85	7 - 33	56 - 91

Sieve analyses were performed for a limited number of char samples. The results are given in Appendix 2. As would be expected, the char samples had a smaller average diameter than the corresponding feed materials. For the NM coal, the average char particle diameter ranged from 270 to 740 microns. Compared to the feed, the char average diameter ranged from 50-90% of the feed average diameter, with an average value of 78%. The lignite was apparently a more friable material, as the char average particle diameter ranged from 170 to 540 microns. The lignite char particle size averaged 54% of the feed particle size.

SUMMARY AND CONCLUSIONS

A devolatilized Kentucky bituminous coal, a New Mexico subbituminous coal, a North Carolina peat, and a Texas lignite were gasified with steam and oxygen in a pilot-scale fluidized bed reactor. The reactor was operated at pressures of 570-840 kPa (80-120 psia), molar steam-to-carbon feed ratios of 0.6 to 1.9, and average bed temperatures of 795-1010°C (1460-1850°F). The coal feed rate ranged from 14 to 33 kg/hr (30-73 lb/hr). All reactor effluent streams were measured and analyzed, enabling performance of mass and energy balances and identification of potentially hazardous species.

Carbon conversions were calculated as the percent of the carbon in the feed coal converted to a gaseous form. The calculated conversions ranged from 16% for a Kentucky char run to 91% for a New Mexico coal run. The Kentucky char was the least reactive of the materials studied, and the North Carolina peat and the Texas lignite were the most reactive.

The effect of the reactor operating variables on carbon conversion and gas production were determined from runs which effectively isolated particular variables. Conversion and gas production both increased with increasing temperature and steam/carbon feed ratios. Increasing the reactor temperature increases the extent of devolatilization and the reaction rates in the fluidized bed. Increases in the steam/carbon ratio promote the steam-carbon reaction, which is the primary gasification reaction. Increases in the solid-phase space time increased both conversion and gas production when the increase was achieved by increasing the bed height. The solid-phase space time can also be increased by lowering the coal feed rate. Increases achieved in this manner give an increased gas production rate, but yield a lower carbon conversion. Insufficient experimental data were obtained to determine the effect of pressure on reactor performance. For the small pressure range covered, 65 to 105 psig, no significant effects were observed.

Gasification of the devolatilized Kentucky coal yielded a gas composed primarily of H_2O, CO, CO_2, H_2, CH_4, N_2, H_2S, and COS. The raw coal feedstocks yielded the same gaseous species plus a wide range of hydrocarbon and sulfur gases. The hydrocarbon species measured were aliphatics (CH_4 to C_4H_{10}) and aromatics (benzene, toluene, xylenes, and ethylbenzene). In addition to H_2S and COS, the sulfur species CS_2, CH_3SH, C_2H_5SH, and thiophene were detected in the effluent gases for the

raw coal feedstocks. The gas compositions, expressed as mole percents on a dry N_2-free basis, showed small differences with the exception of the sulfur gases. The differences in sulfur gas concentrations were generally proportional to the sulfur contents of the feed coals.

The gas production rates were normalized to give production rates in grams per kilogram of dry ash-free coal fed to the reactor. Sulfur gas rates were proportional to the sulfur content of the feed coals. The distribution of gaseous sulfur species was similar for the New Mexico coal and Texas lignite; however, the North Carolina peat gave a higher proportion of methyl mercaptan and thiophene. Comparison of the normalized species rates showed similar values for aliphatic species with the exception of methane. The New Mexico coal yielded methane rates greater than those obtained from the North Carolina peat and Texas lignite. Methane yields from the Kentucky coal were much lower than those from the raw coal feedstocks, undoubtedly a result of pretreatment. Similar aromatic hydrocarbon production rates were obtained from the raw coals.

Wastewater samples were taken from the sample train and subjected to several standard analyses. The wastewater from gasification of the Kentucky char was relatively clean, again a consequence of the pretreatment the coal had undergone. The raw coal feedstocks yielded a much filthier wastewater. The concentrations and production rates of the species in the wastewater from gasification of the raw coals were nearly the same with the exception of dissolved carbon (carbon, COD, TOC, and TVC). Peat yielded a wastewater with much higher levels of dissolved carbon than either the subbituminous coal or lignite, for which the levels were roughly the same. No correlation of wastewater species production rates with reactor operating conditions were observed.

Tar production rates were calculated from cold trap measurements. The devolatilized Kentucky coal yielded no measureable tar, while the other feedstocks gave tar production rates of 0.05 to 0.14 kg/kg daf coal fed. The tar production rate was greatest for the New Mexico subbituminous coal, for which the tar rate ranged from 1.5 to 2.2 kg/hr (0.10 to 0.14 kg/kg of the dry ash-free feed). Tar production rates for the North Carolina peat and the Texas lignite were similar, ranging from 0.9 to 1.5 kg/hr (0.05 to 0.10 kg/kg of the daf feed). The tar contained a significant fraction of the carbon in the feed coal. Based on estimated tar compositions, the carbon lost in the tar ranged from 7 to 16% of the feed carbon. This value can be reduced significantly by intra-bed feeding of the coal.

Particulates entrained with the reactor exit gas were collected in a cyclone separator, and particulates escaping the cyclone were trapped in the sample train cold trap. Cyclone efficiencies were calculated to be as great as 96%. However, for several runs the cyclone did not function, probably due to tar condensation. The total particulate

collection rates ranged from 1.0 to 3.7 kg/hr, or 4.7 to 16.4% of the coal feed rate. The carbon in the particulates represented from 3 to 17% of the feed carbon, indicating the desirability of recycle of particulates to the reactor. The elutriation rates, expressed as percents of the coal feed rate, were greatest for the Texas lignite. The feed particle size distributions were similar for all of the feeds; the greater elutriation rates observed for lignite thus suggest that it is a more friable material than the other feeds.

Samples of the spent char were taken from two locations following completion of a run -- the char receiver and the bed. Analyses of char samples from the two locations yielded nearly identical results. This indicates that the solids in the fluidized bed may be assumed to be well-mixed. As would be expected, the char samples generally showed a greater ash content and a lower carbon content than the feed materials. The ash content was as high as 91% for a Texas lignite run.

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